T28

(282)

SSAB

Santa Cruz

Volume 3

Memorandum re: Tests

P.M. beträffande Santa Cruz - försöken

Sammandrag av tidigare försök

Santa Cruz - försöken startades försommaren 1955 av SSAB och Husky Oil Co. i avsikt att på basis av erfarenheter från Kvarntorp utforma LINS-metoden för oljeutvinning ur tjärsand. De första försöksserierna, omfattande ett antal enhåls- och sjuhålsförsök, bekräftade, att det var möjligt att erhålla olja på detta sätt samt gav underlag för beräkningar över hålavstånd, brännareffekter, upphettningstider etc.

Därpå följde ett antal försök för studium av olika brännartyper. Ett utförande, som ansågs vara användbart, beslöts provas i en större försöksenhet, omfattande 100 värmehål. Under loppet av 1957, då driften av detta första hundrahålsförsök pågick, konstaterades emellertid, att värmerören och brännarna ej uthärdade de högre temperaturer i berget, som nu uppnåddes. (Det hade av fysikaliska orsaker ej varit möjligt att i de tidigare, mindre försöken prova utrustningen vid så hög temperatur.)

Hundrahålsförsöket måste avbrytas, och mera arbete nedläggas på att ytterligare förbättra brännarna. Speciellt gällde det att eliminera risken för lokala värmekoncentrationer. En ny brännartyp såg dagens ljus sommaren 1957. Mot slutet av samma år beslöts, att ett nytt hundrahålsförsök skulle startas. Detta försök (kallat "L 9") pågick under hela 1958 och avslutades programenligt i början av 1959. Programmet för Santa Cruz - försöken var därmed genomgånget, och verksamheten där upphörde på försommaren 1959.

Ändamålet med försök L 9

För bedömning av metodens ekonomi är det väsentligt att veta vilka utbyten av olja och gas, som kan påräknas från en anläggning i större skala. Hundrahålsförsöket L 9 planerades så, att resultat i detta avseende skulle erhållas. Dessutom var det naturligtvis angeläget att erhålla ytterligare erfarenheter från längre tids drift av de nya brännarna. Vidare skulle den lämpligaste anordningen av uttagshålen för producerad gas och olja ytterligare studeras.

Försökets utförande

Försöket förlad s i omedelbar närhet av tidigare försöksenheter. Ett stort

antal borrkärne-analyser visade, att tjärsanden var relativt homogen mellan 10 och 45 fots djup med en genomsnittlig tjärhalt av 184 kg/m^3 berg.

Etthundra brännarhål borrades i ett triangulärt mönster med ett hålavstånd av 3,05 meter (10 fot). Hela försöksfältets yta var 690 m^2. Det uppvärmda tjärsandslagret var cirka 12 meter tjockt. (Som diskuteras nedan, kan utbytesberäkningar ej göras på basis av dessa mått, på grund av den kalla omgivningens kylande inverkan på de yttre delarna av försöksfältet.)

Samtliga brännarhål var så utrustade, att producerad gas och olja kunde uttagas genom samma borrhål (koncentriskt med eller vid sidan av brännarens ytterrör). Vidare borrades 23 separata gashål, 22 temperaturmätningshål och 14 hål för grundvattenpumpning.

I brännarhålen nedsattes brännare av den konstruktion, som visas å bilaga 1. Brännaren bestod sålunda av ett ⅛" eller 1/4" bränsle-nedledningsrör, en brännarkona och ett cirka 5 meter långt 1" brännarrör, alltsammans nedsatt i ett cirka 16 meter långt, nedtill slutet 2½" ytterrör. För centrering i ytterröret var brännarröret försett med påsvetsade styrfenor. Ytterrören var gjorda av en stållegering, innehållande cirka 5 % Cr, 1,4 % Si och 0,5 % Mo. (Trots att det ansågs tämligen säkert, att olegerat material skulle ha kunnat användas, beslöts att i detta försök eliminera varje risk för rörhaverier genom användning av ett legerat material.)

Brännarrören var tillverkade av dels 25 - 20, dels 18 - 8 Cr - Ni - stål. Konorna var av 25 - 12 - stål och nedledningsrören av olegerat stål, utom de nedersta 2 meterna, som var av 18 - 8 - stål.

Mellanrummet mellan brännarrör och ytterrör var fyllt med sand i sådan mängd och kornstorlek, att en jämn fördelning av svävande sandkorn utefter brännarens längd erhölls då brännaren var tänd. Med jämna mellanrum påfylldes mindre mängder ny sand som ersättning för den sand, som nötts ut och i form av fint stoft bortgått med rökgaserna.

Brännarna var i drift från februari 1958 till januari 1959 (i 8.057 timmar) med en tillförd effekt, som under största delen av försökstiden var cirka 7.000 kcal/timma. Mot försökets slut nedreglerades effekten något. Totalt inmatades i hela fältet 4.900.000 Mcal värme. Som bränsle användes propan. Fältets egen gasproduktion motsvarade en värmemängd av 1.130.000 Mcal.

Drifterfarenheter av brännarna

Under försöket havererade fem brännar-ytterrör. I fyra av dessa fall var orsaken sticklågor (från brustna nedledningsrör eller andra orsaker), och i ett fall hade för litet sand fyllts i röret med ojämn värmefördelning som följd. I samband med försökets avbrytande brast ett rör, sannolikt beroende på förskjutningar i berget.

sålund!

Tre brännarkonor brändes sönder av olika orsaker. Inget brännarrör havererade.

Sammanlagt var brännarna ur drift under endast 2,95 % av försökstiden. Därav förorsakade:

strömavbrott	0,49 %
underhåll på bränsleledningar och dylikt	0,59 %
brännarhaverier	0,83 %
inspektion och underhåll av brännare	0,24 %
diverse orsaker utanför brännarna	0,80 %
	2,95 %

Produktionen av olja och gas

Produktionen uttogs praktiskt taget helt genom de uttag, som var anordnade vid brännarna. Tjärsandens genomsläpplighet var för låg för att tillåta någon avsevärd mängd produkter att uttränga genom de separata uttagshål, som provades.

Under försökets första del uppträdde ett stort antal igenpluggningar av ledningarna av opyrolyserad tjära, som av vattenångan ryckts med upp i uttagshålen. Så småningom - när bergets temperatur stigit mera - försvann dessa svårigheter. En tunn olja erhölls, som lätt strömmade genom ledningarna. För hävande av tendenser till emulsionsbildning mellan oljan och pyrolysvattnet doserades minimala kvantiteter av ett emulsionsbrytande preparat till blandningen.

Under framför allt försökets senare hälft började läckage uppstå i marken, omkring och mellan brännarna. Tjära, vattenånga, oljeångor och gas trängde upp. Tätning med cementvälling försöktes men utan större framgång. Orsaken till läckagen antogs vara, att fältets "overburden" var otillräcklig. (Den bestod

här av en halvmeter matjord och cirka tre meter tjärsand.) Sedan uppvärmningszonen flyttats två meter djupare ned (genom sänkning av brännarna), upphörde markläckaget i stort sett.

Sammanlagt producerades:

423 m^3 råolja
128.000 Nm^3 rågas
1.470 m^3 pyrolysvatten

Analys av råoljan (generalprov):

	Santa Cruz - oljan	För jämförelse: råolja från Ljungströmsanl. i Kvarntorp
spec. vikt	0,888	0,881
svavelhalt	2,15 %	1,37 %
kvävehalt	0,38 %	0,35 %
förkoksningsrest	0,11 %	0,48 %
ASTM-dest.:		
5 % överdest.	128° C	106° C
10 %	146	127
20 %	186	149
30 %	229	175
40 %	268	197
50 %	299	218
70 %	338	267
95 %	407	-

Analys av rågasen (generalprov):

	Santa Cruz - gas	För jämförelse: rågas från Ljungströmsanl. i Kvarntorp
H_2	39,6 %	19,9 %
H_2S	9,2	25,0
CO_2	2,3	1,4
$N_2 + CO$	0,7	5,5
CH_4	28,6	37,8
C_2	7,6	8,2
C_3	4,6	3,6
C_4	4,3	1,5

Analys av rågasen (generalprov) fortsättning:

C_5	2,4 %	1,9 %
C_6	0,7	–
	100,0 %	100,0 %
Värmevärde, eff.	8,1 Mcal/Nm^3	8,65 Mcal/Nm^3

Utbyten

Avsikten var från början att uppmäta fältets totala produktion och slå ut den på den totala, uppvärmda bergvolymen, korrigerad för randförluster m.m. Under bearbetningens gång visade det sig emellertid, att osäkerheten skulle bli stor vid ett dylikt förfarande. Mera tillförlitliga resultat borde kunna erhållas, om man utvalde några smärre delar av fältet, belägna i så "temperaturjämna" och "tryckjämna" områden, att det kunde anses, att ingen nettoströmning av gaser eller vätskor skett till eller från dessa ställen. Tre provytor utvaldes. Ett stort antal borrkärnor från dessa (och övriga) delar av fältet upptogs efter försökets slut och analyserades.

Provyta nr	1	2	3
Ursprungligt tjär-innehåll	1.160 kg	1.310 kg	960 kg
Utvunnen olja	500 kg = 43 %	666 kg = 51 %	510 kg = 53 %
Utvunnen gas	150 kg = 13 %	198 kg = 15 %	153 kg = 16 %
Totalt utbyte av kol-väten	650 kg = 56 %	864 kg = 66 %	663 kg = 69 %

Beträffande provyta nr 1 må anmärkas att ett av dess gashål var pluggat av tjära under en viss del av försökstiden. Det är sålunda sannolikt, att det lägre utbytet från denna provyta förklaras därav.

Till jämförelse anföres resultatet av en vanlig Fischer-analys på tjärsand:

Ur ett prov, innehållande		12,56 gram tjära
erhölls:	olja	8,44 gram = 67 %
	gas	0,79 gram = 6,3 %
	totalt utbyte av kolväten	9,23 gram = 73,3 %

Provyta nr 2 gav sålunda 76 % av Fischer-utbytet och
provyta nr 3 gav sålunda 80 % av Fischer-utbytet
räknat på enbart oljan (90 respektive 94 % räknat på oljan + gasen).

Värmebalanser

Av det tillförda värmet användes blott en del för uppvärmning av själva tjärsanden. En del bortgick med utgående rökgaser, en del spriddes via overburden till atmosfären, en del spriddes horisontellt via ledning eller via strömmande grundvatten till omgivande tjärsand och en del spriddes till underliggande berglager. På basis av temperaturmätningar och teoretiska beräkningar uppställdes följande försök till balans för hundrahålsförsöket:

1)	för upphettning och pyrolys av tjärsanden	$680 \cdot 10^3$ Mcal	=	14 %
2)	för bortkokning av grundvatten	780	-"-	= 16
3)	förluster till underliggande lager	920	-"-	= 18
4)	" " sidobelägna "	1.340	-"-	= 28
5)	" " overburden och atmosfären	540	-"-	= 11
6)	" genom utg. rökgas	640	-"-	= 13
7)	totalt tillfört värme	$4.900 \cdot 10^3$ Mcal	=	100 %

Det må anmärkas, att balansen är relativt osäker. Dock framgår det med all tydlighet, att i ett stort fält med ett tjockare tjärsandlager och effektivare grundvattenbortpumpning den relativa värmetillförseln kan reduceras till under hälften av ovanstående siffror.

Den producerade gasmängden motsvarar 23 % av det tillförda värmet. För att anläggningen skall bli självförsörjande med gasbränsle behövs tydligen, utom att ovanstående villkor är uppfyllda, också att tjärhalten är högre än i Santa Cruz-fyndigheten.

Kvarstående problem

Vid slutsammanträde mellan Husky Oils, Union Oils och SSAB:s representanter i Santa Cruz den 20 maj 1959 genomgicks ovan relaterade försöksresultat. Husky Oils och Union Oils representanter förklarade sig vara mycket tillfredsställda med utgången av Santa Cruz - projektet. Speciellt uppmärksammade man de goda utbytessiffrorna, den höga kvaliteten på oljan och den mycket goda driftsäkerheten på brännarna. Man förklarade sig under den närmaste tiden önska göra ekonomiska kalkyler på basis av försöksresultaten. Vidare ansågs det aktuellt att börja planera en halvstor anläggning i Athabasca-området på en rikare och mäktigare tjärsand än Santa Cruz - tjärsanden. Husky Oil skulle under hösten utarbeta och tillsända Union Oil och SSAB ett förslag till en dylik anläggning.

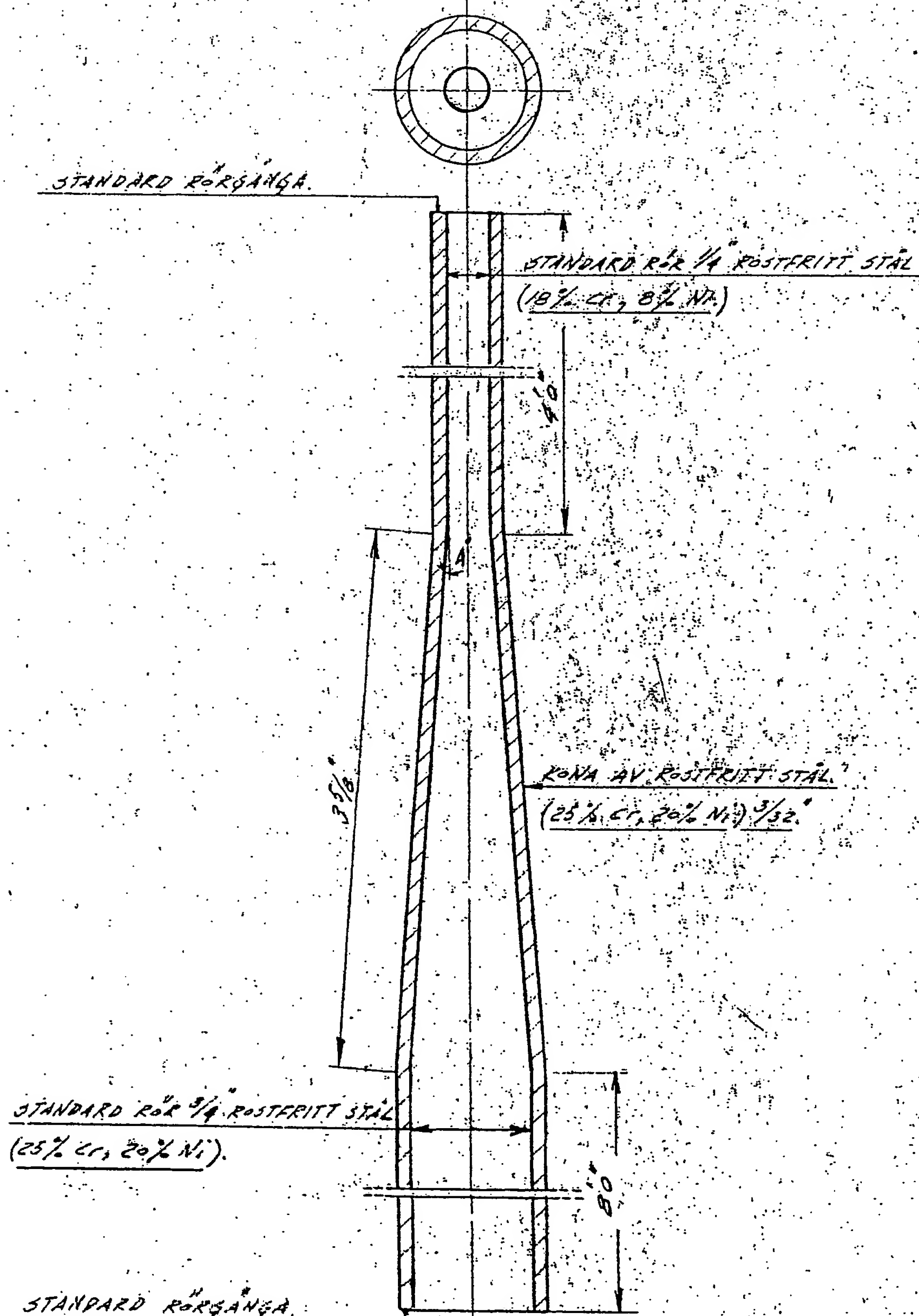
Närkes Kvarntorp den 21 augusti 1959

Gösta Salomonsson

Överingenjör

1 bilaga

Utl. den av till



B A Antal	Detalj nr	Benämning	Beteckning Dimension	Material	Vikt	Anmärkning

Konstr.	Ritad	Kontr.	Godkänd	Stand.	Datum	Skala	Ersätter	Ersatt av
	Bl.				20.8.1955			

KVARN TORP

SVENSKA SKIFFEROLJE AB

LINS BRÄNNARRÖR

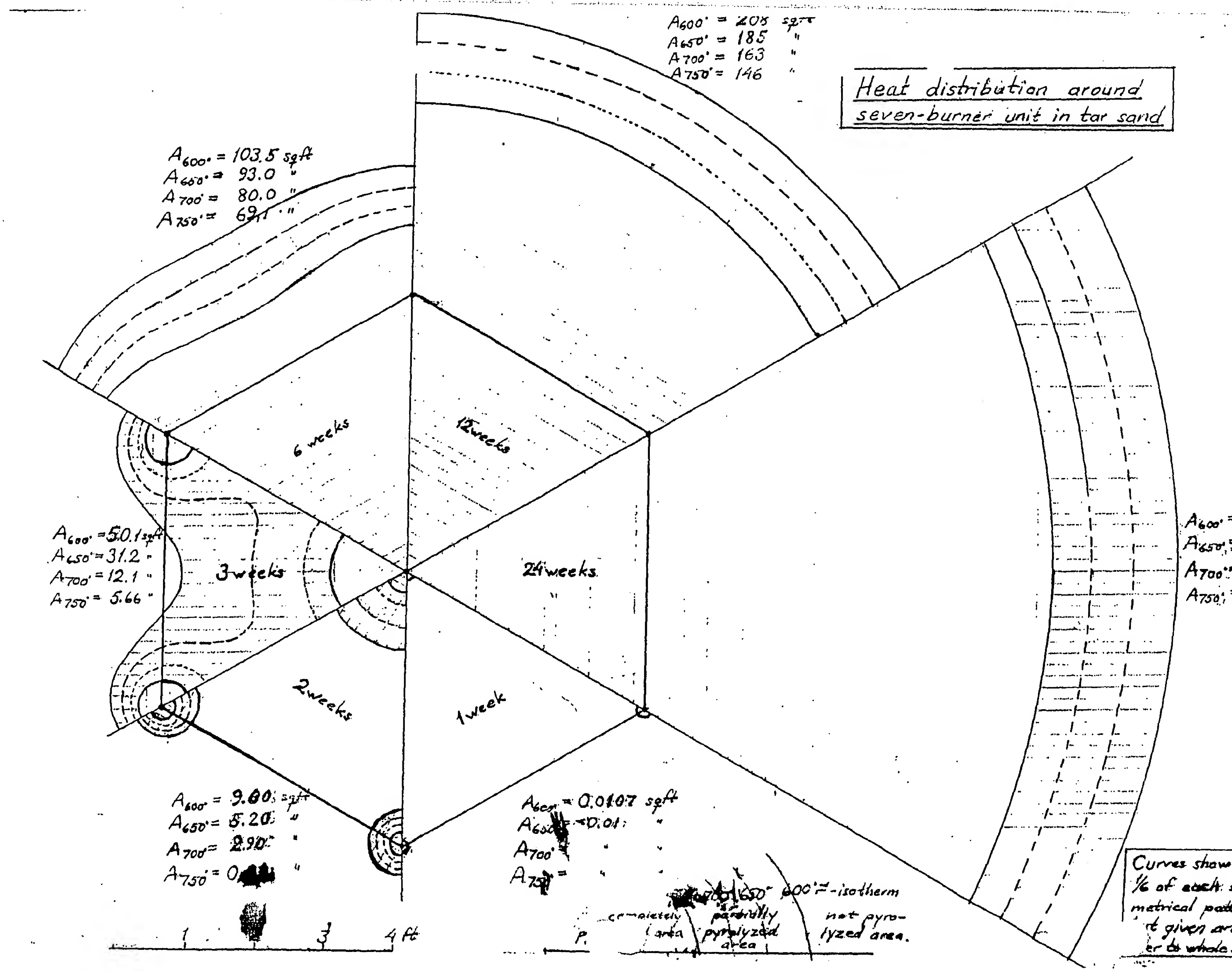
Field volume (effectively heated): 77 × 62.7 × 40 = 4830 × 40 = 193.000 cuft

Tar content: 8.5 % by weight = 5280 barrels

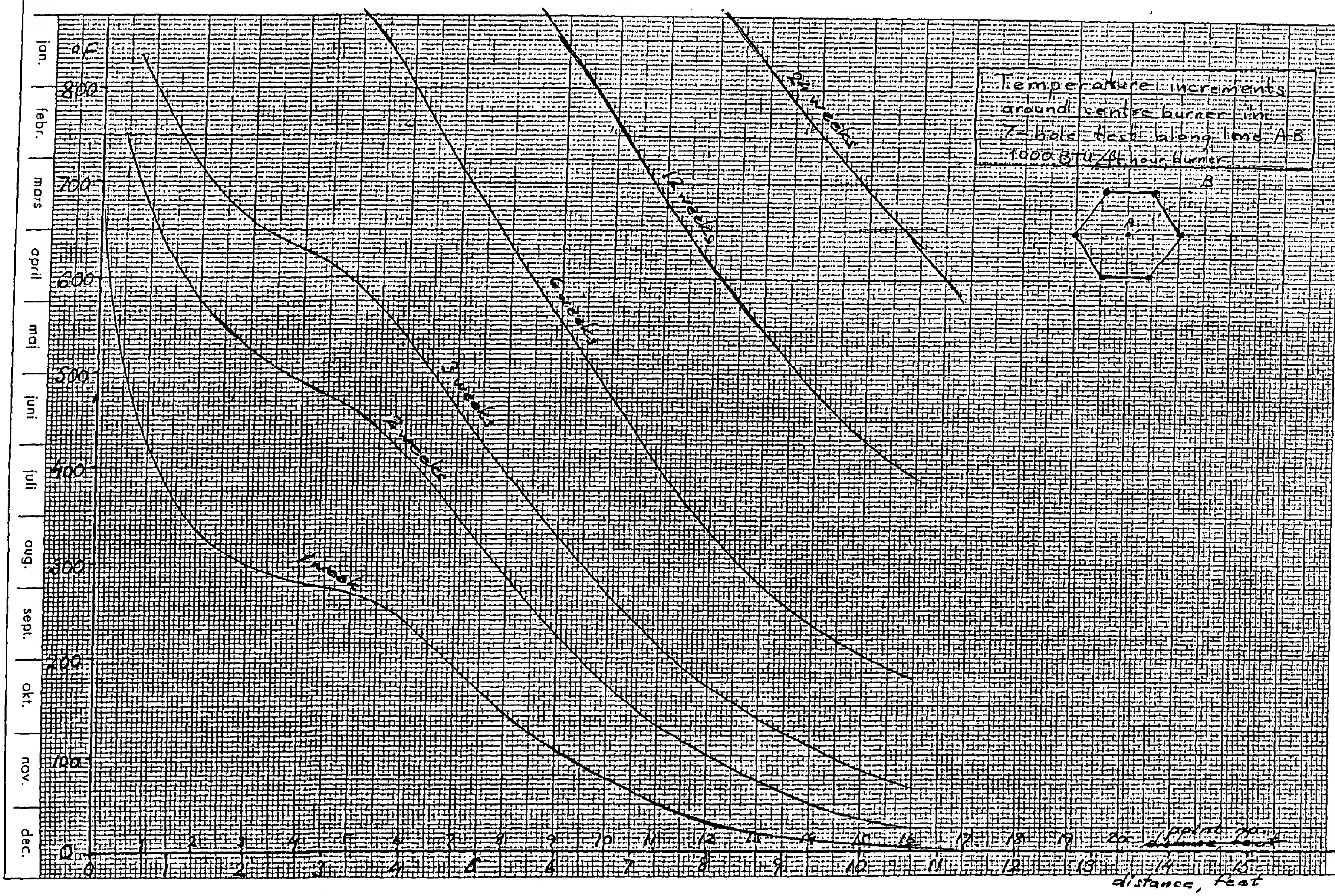
Total amount of heat to be supplied = 17.000 · 10^6 BTU.

Rows 1-4 started ~~Jan~~ Feb. 25, Rows 5-8 Feb. 19, Rows 9-10 March 26.

Week	Week no.	Heat input, BTU week	Heat input, BTU cum.	Production Oil, bbl week	Production Oil, bbl cum.	Production Gas, week	Production Gas, cum.	Production Water bbl week	Gas/oil cuft/bbl week
before 19.5.	169	—	4204	—	2.4	10^3 cuft	10^6 cuft	—	( )
19.5 – 25.5	170	438		18.1		—	—	96	
26.5 – 1.6.	171	450		29.2		—	—	44	
2.6. – 8.6.	172	464		19.7		10.700		310	
9.6. – 15.6.	173	450		28.2		7.700		278	
16.6. – 22.6.	174	453	6459	14.8	112.4	7.700		297	
23.6. – 29.6.	175	427		16.9		16.200		208	
30.6. – 6.7.	176	447		25.4		18.900		285	
7.7. – 13.7	177	446		25.6		27.100		245	
14.7. – 20.7.	178	468		26.5		27.000		242	
21.7. – 27.7.	179	469	8706	34.2	241.0	33.900		216	
28.7. – 3.8.	180	463	~~8706~~	39.2	~~[illegible]~~	34.800		199	
4.8 – 10.8.	181	452	9578	38.8	311	54.600	235	164	
11.8. – 17.8.	~~184~~ 182	455	10.032	41.7	353	58.800	293	187 ~~185~~	
18.8. – 24.8.	~~185~~ 183	467	10.499	50.0	403	77.000	370	182	
25.8. – 31.8.	~~183~~ 184	467	10.966	57.4	460	91.500	462	182	
1.9 – 7.9.	~~184~~ 185	454	11.421	67.8	528	106.400	568	188	
8.9. – 14.9	186	457	11.872	78.7	607	112.000	680	156	
15.9. – 21.9.	187	447	12.319	85.2	692	124.800	805	152	
22.9. – 28.9.	188	446	12.765	98.3	790	147.100	952	132	
29.9 – 5.10	189	423	13.188	115	905	170 · 10^3	1122	90	
6.10. – 12.10	190	422	13.610	130	1035	201	1324	116	
13.10 – 19.10	191	422	14.032	132	1176	231	1555	150	
20.10. – 26.10.	192	426	14.457	135	1311	231	1786	168	
27.10 – 2.11.	193	394	14.853	127	1437	217	2002	158	
3.11 – 9.11.	194	391	15.244	118	1555	220	2222	159	
10.11. – 16.11.	195	389	15.632	126	1680	223	2445	168	
17.11. – 23.11	~~195~~ 196	383	16.015	105	1785	223	2668	153	
24.11. – 30.11.	~~196~~ 197	415	16.430	100	1885	229	2898	113	
1.12. – 7.12.	~~197~~ 198	410	16.840	95	1980	227	3124	126	
8.12. – 14.12.	~~198~~ 199	387	17.228	89	2068	215	3339	155	
15.12. – 21.12.	~~199~~ 200	392	17.619	83	2157	187	3526	132	
22.12. – 28.12.	~~200~~ 201	389	18.007	73	2224	177	3704	164	
29.12 – 4.1.59	202	385	18.393	67	2291	167	3871	180	
5.1 – 11.1.	203	382	18.775	93	2384	190	4060	278	
12.1. – 18.1	204								
19.1. – 25.1	205	352	19.495	47	2488	113	4311	225	
26.1. – 31.1.	206		19.545		2577		4375	[illegible] 7.906	
1.2. – 8.2.	207				2548		4429	[illegible] 8153	
	208								

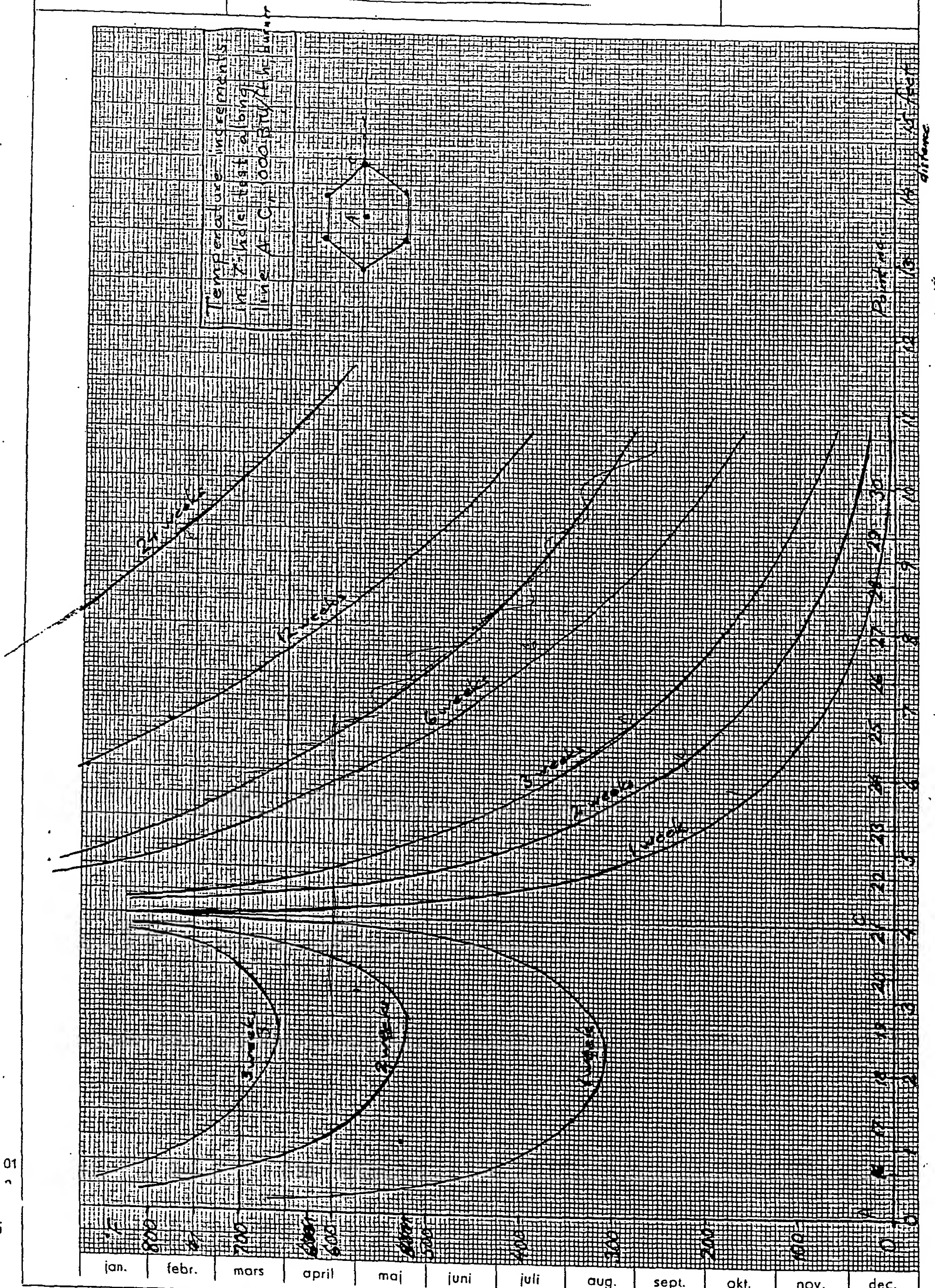


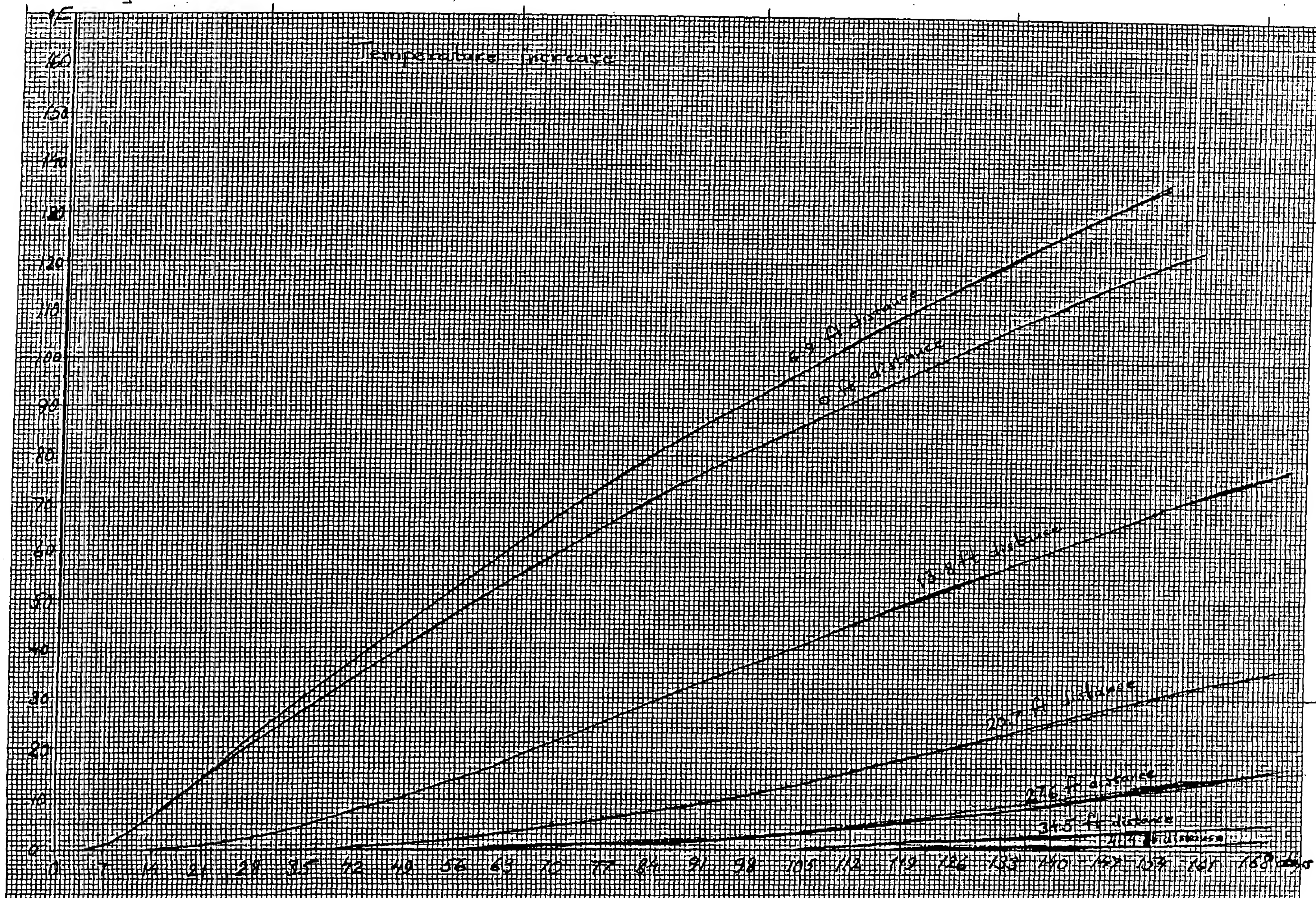
Heat distribution around seven-burner unit in tar sand
A600' = 208 sq ft
A650' = 185
A700' = 163
A750' = 146
A600' = 103.5 sq ft
A650' = 93.0
A700' = 80.0
A600' = 50.1 sq ft
A650' = 31.2
A700' = 12.1
A750' = 5.66
A600' = 9.00 sq ft
A650' = 5.20
A700' = 2.90
A600' = 0.0107 sq ft
A650' = 0.01
A700'
6 weeks
12 weeks
3 weeks
24 weeks
2 weeks
1 week
1
3
4 ft
600°F-isotherm
partially pyrolyzed area
not pyrolyzed area
Curves show

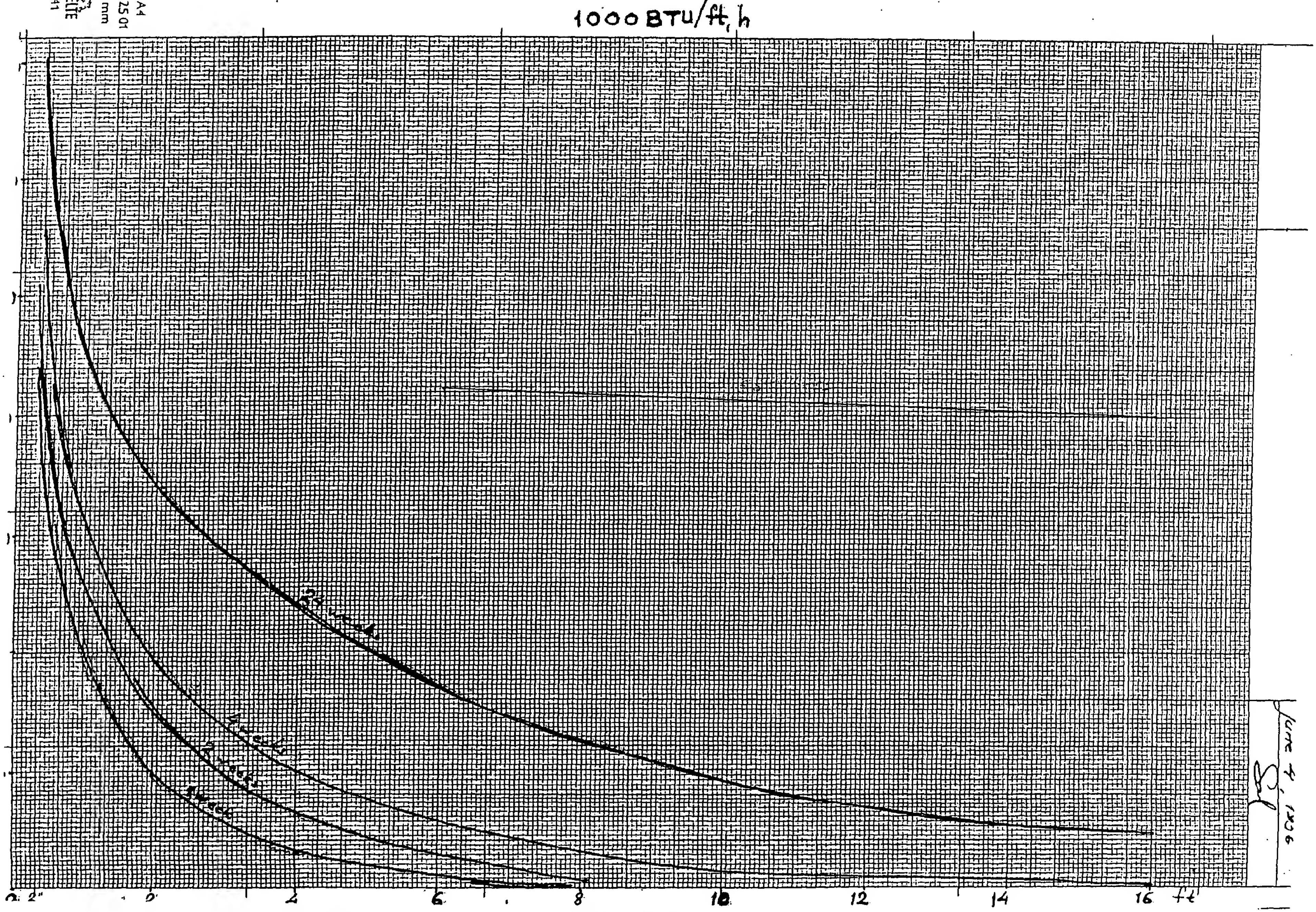


744 A4
73 25 01
x mm
ESSELTE
4474

line 16 - 20.




°C
Temperature increase
160
150
140
130
120
110
100
90
80
70
60
50
40
30
20
10
0
0
7
14
21
28
35
42
49
56
63
70
77
84
91
98
105
112
119
126
133
140
147
154
161
168
days
6.9 ft distance
0 ft distance
13.8 ft distance
20.7 ft distance
27.6 ft distance
34.5 ft distance
41.4 ft distance



514 A4
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1×1 mm
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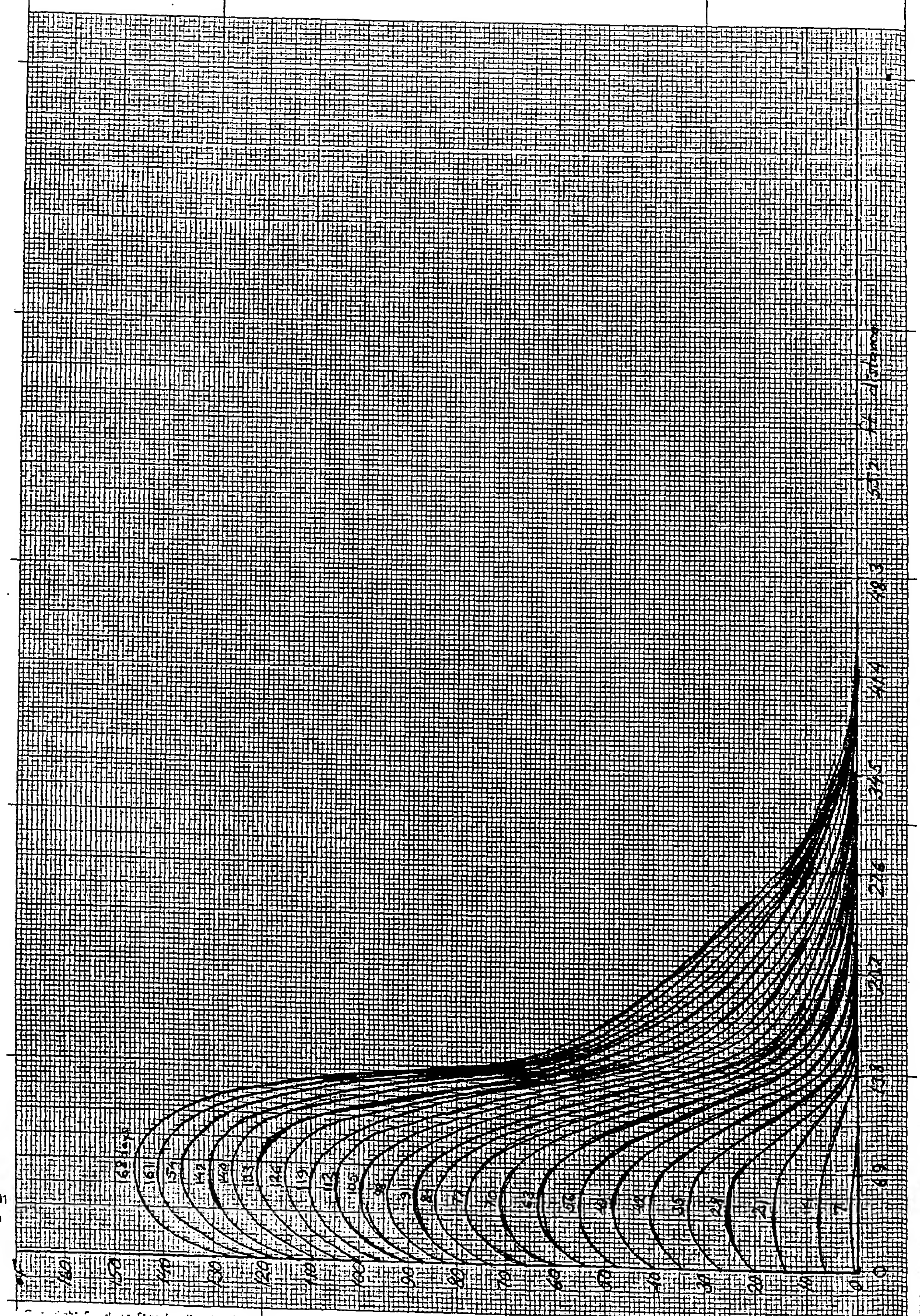


0
10
20
30
40
50
60
70
80
90
100
110
120
130
140
150
160
6.9
13.8
20.7
27.6
34.5
41.4
48.3
55.2
distance
168 days

Sal

Program I: uniform ...

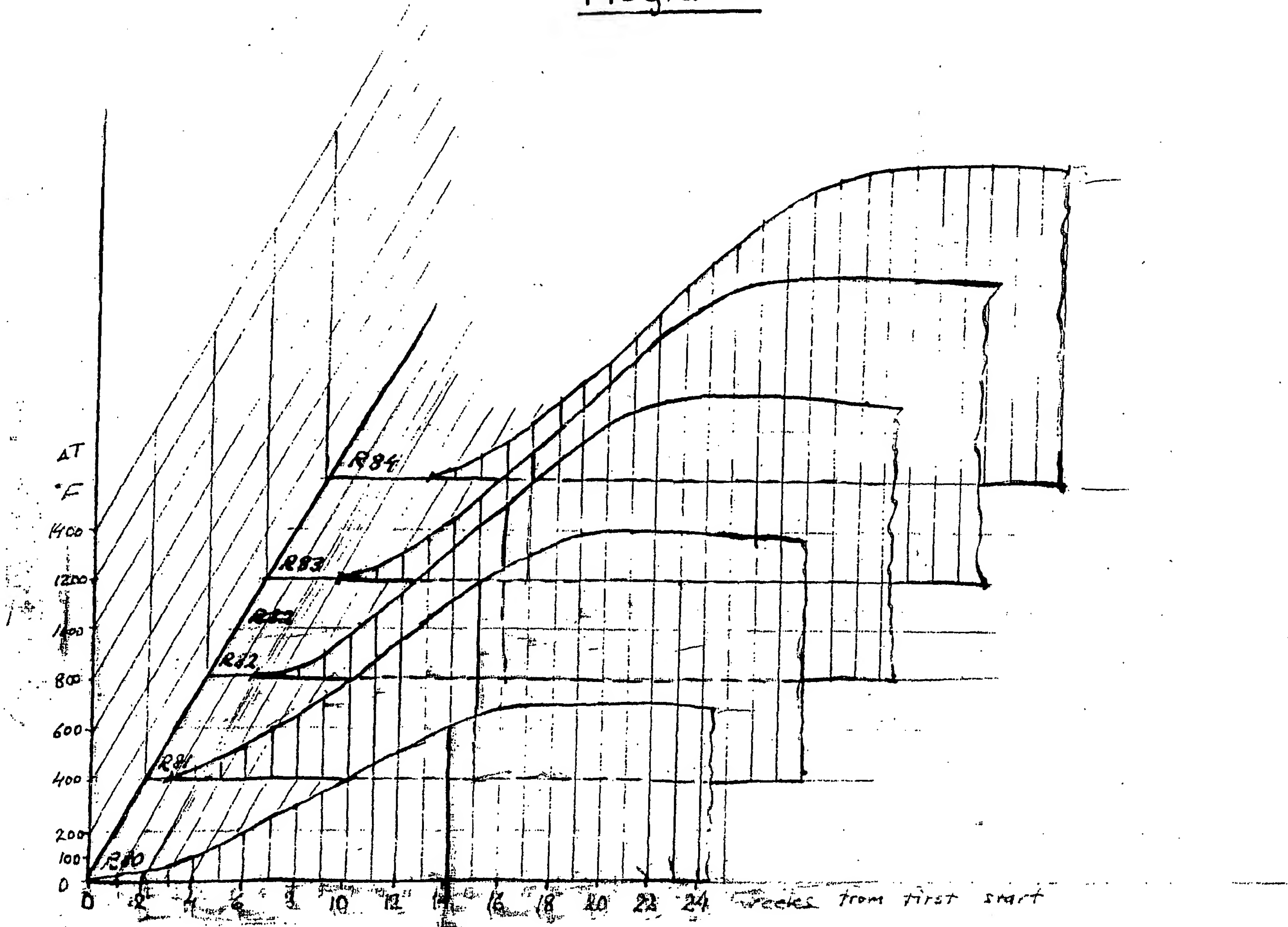
row started each week, shut off after 14 weeks.

days from ~~first shut~~	Row 80			Row 81			Row 82			Row 83			Row 84		
Start of actual row	ΔT_i	ΔT_o	$\Sigma\Delta T$	ΔT_i	ΔT_o	$\Sigma\Delta T$	ΔT_1	ΔT_o	$\Sigma\Delta T$	ΔT_1	ΔT_o	$\Sigma\Delta T$	ΔT_1	ΔT_o	$\Sigma\Delta T$
28	533 ~~462~~	108	661		198			221			226			227	
35	571 ~~583~~	162	733		276			308			315			317	
42	583	203	786		338			378			388			391	
49	594 ~~[illegible]~~	250	844		406			458			474			478	
56	605 ~~571~~	303	908		486			550			570			575	
63	614 ~~583~~	353	967		557			632			655			662	
70	621 ~~594~~	408	1029		633			720			749			758	
77	628 ~~605~~	462	1090		705			804			840			850	
84	636 ~~614~~	579	1153		777			888			930			944	
91	641 ~~621~~	570	1211		846			969			1016 ~~977~~			1033 ~~994~~	
98	646 ~~628~~	614	1260		905 ~~908~~			1027 ~~1043~~			1092 ~~1096~~			1113 ~~1117~~	
105	189 ~~[illegible]~~ ~~636~~	664	860 ~~[illegible]~~		953			1097			1156			1183	
112	152 ~~[illegible]~~ ~~641~~	691	843 ~~[illegible]~~		974			1125			1192			1222	
119	126 ~~659~~ ~~646~~	705	831 ~~[illegible]~~		983			1139			1211			1243	
126	111 ~~664~~ ~~657~~	710	821 ~~[illegible]~~		981 ~~985~~			1139 ~~1142~~			1217 ~~1220~~			1252 ~~1255~~	
133	97 ~~668~~ ~~655~~	714 ~~717~~	811 ~~[illegible]~~		976 ~~981~~			1137 ~~1144~~			1220 ~~1227~~			1257 ~~1264~~	
140	88 ~~671~~ ~~657~~	718	806		975 ~~977~~			1140 ~~1144~~			1228 ~~1232~~			1268 ~~1272~~	
147	81 ~~[illegible]~~ ~~664~~	718 ~~[illegible]~~	799		967 ~~[illegible]~~			1131 ~~[illegible]~~			1220 ~~1240~~			1262 ~~1282~~	
154	74 ~~[illegible]~~ ~~[illegible]~~	724	798		964			1128			1221			1265	
161	68 ~~682~~ ~~674~~	724	792		958			1123			—			—	
168	64 ~~685~~ ~~673~~	717	781		945			—			—			—	
~~175~~															

ΔT_o = temp. increase 2" from burner center caused by all outside burners.
ΔT_i = temp. increase 2" from burner center, caused by inside burner. $\Sigma\Delta T = \Delta T_i + \Delta T_o$

ΔT °F 1400 1200 1000 800 600 400 200 100 0

Program 1.



ΔT
°F
1400
1200
1000
800
600
400
200
100
0
R80
R81
R82
R83
R84
0
2
4
6
8
10
12
14
16
18
20
22
24
weeks from first start

Type A (burner-tube, no jet, no hood)

3/4" – burners

#	length	input BTU/h	hours	total input 10^6 BTU	cone mat	observ.	casing mat	observ.	remarks
L2	27'	29,000	1232	37	w25/20	/	18/8	—	—
L2	27'-20'	29,000	332	10	w25/20	—	18/8	X	—
L21	12'	16-21,000	1257	25	w25/20	—	iron ~~X~~	(X)	burner tube split
L3	27'	30.000	6164	189	w25/20	(X)	18/8	—	cone re-paired
L6	17'	25.000	1329	33	c25/12	—	18/8	—	—
~~L6~~	~~17'~~	~~23.000~~	~~439~~	~~10~~	~~c25/12~~				

General observations: ~~no damage ... (except when burner tube split)~~ Small corrosion on w25/20-cones, at high ~~input~~ temperatures (after long burner runs).

1" – burners

#	length	input BTU/h	hours	total input 10^6 BTU	cone mat	obs.	casing mat	obs.	remarks
L4	27.5'	~40.000	495	20	w25/20	/	18/8	—	—
L4	27.5'	39.000	795	31	w25/20	—	18/8	—	—
L5	27.5'	42.000	167	7	w25/20	X	18/8	(X)	—
L5	27.5'	39.000	335	13	w25/20	—	18/8	X	—
L5	27.5'	39.000	182	7	w25/20	—	18/8	X	—

General observations: ~~18/8 casing damaged, when cone of w25/12 material, not when cone of w25/20. Cone w25/12 cone burnt off~~

3/4"-burners

#	length	input BTU/h	hours	total input 10^6 BTU	cone mat	cone obs	casing mat	casing obs	remarks
L6	17'	23.000	439	10	c25/12	—	18/8	—	burner tube split

1"-burners

#	length	input BTU/h	hours	total input 10^6 BTU	cone mat	cone obs	casing mat	casing obs	remarks
L4	27.5'	34.000	2739	100	w25/20	—	18/8	—	—
L5	27.5'	29.000	876	26	w25/12	—	18/8	X	casing off in bottom
L51	20'	40.000	78	3	c25/12	—	iron	—	—
L52	20'	40.000	117	5	c25/12	X	iron	X	burner tube off
L63	20'	40.000	660	26	c25/12	—	iron	X	—
L71B3	10'	20.000	542	11	Kanthal	?	18/8 iron	(X)	casing off because [illegible]?

General observations: high input of heat per hour is more fatal to casing than a high total amount of supplied heat.

Type C (burner-tube, jet, cone, hood)

3/4"-burners (1 3/8" hoods)

	#	length	input BTU/h	hours	total input 10^6 BTU	cone mat	cone obs	casing mat	casing obs	remarks
(3 burners)	L7B2	21'	20-25.000	425	9	w25/20	X	18/8	—	hood corr.
(1 burner)	L7B3	21'	20-25.000	386	8	w25/20	—	18/8	—	burner tube split
(4 burners)	L7B1	21'	20-25.000	445	9	w25/20	—	18/8	—	—
(7 burners)	L7	21'	17-20.000	270	5	w25/20	—	18/8	—	—

1"-burners (2"-hoods)

#	length	input BTU/h	hours	total input 10^6 BTU	cone mat	cone obs	casing mat	casing obs	remarks
L71B4	10'	20.000	691	14	c25/12	X	iron	(X)	casing off at welding seam
L71B5	10'	20.000	541	11	c25/12	(?X)	iron	X	burner stuck
L71B6	10'	20.000	235	5	c25/12	—	iron	X	burner stuck
L31	20'	40.000	115	5	c25/12		iron		burner stuck
L42	20'	40.000	110	4	c25/12		iron		burner stuck

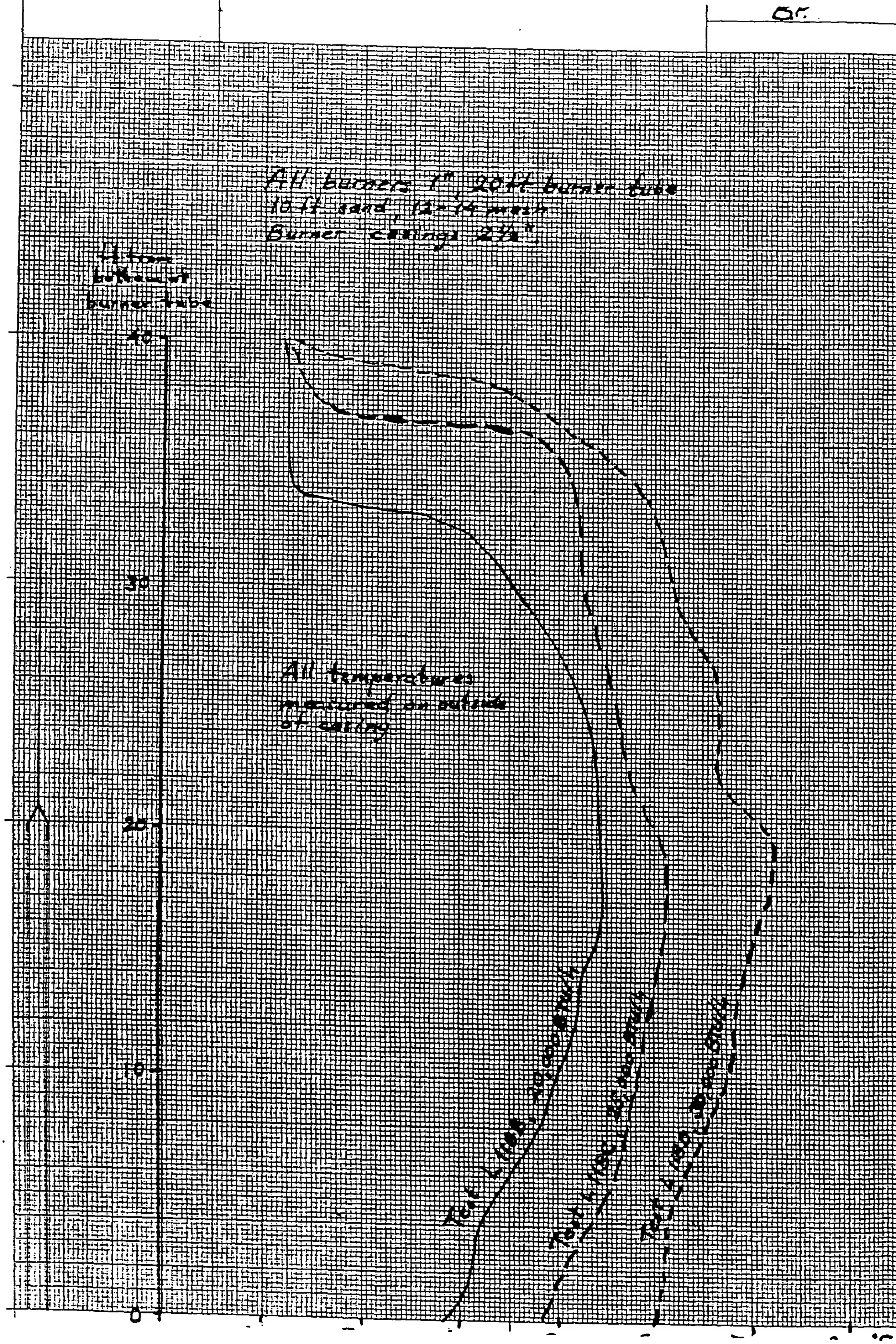
[1" - burner] (1¾" and 2" - hoods)

L41	20'	40.000	161	6	c25/12	—	iron X	~~burner tube badly damaged~~

Type E (short burner tube, jet, one long hood)

[1" - burners]

	B / H							
L4A	15'/28'	34.000	810	28	c25/12	—	18/8 —	burner tube badly damaged
L61	10'/21.5'	20.000	1026	21	c25/12	—	18/8 —	—
L71B1	10'/21.5'	20.000	735	15	w25/20	X	18/8 —	—
L71B1	10'/19.5'	20.000	426	9	Kanthal	—	18/8 X	scaling off because of rock strain
L71B2	10'/21.5'	20.000	641	13	w25/20	X	18/8 X	cone off at welding seam
L71B3	10'/21.5'	20.000	696	14	w25/20	X	18/8 X	—
L71B3	10'/21.5'	20.000	460	9	Fernox	/	18/8 X	—
L71B4	10'/21.5'	20.000	671	14	w25/20	X	18/8 —	cone off at welding seam
L71B4	10'/21.5'	20.000	453	9	c25/12	—	18/8 —	—
L71B5	10'/21.5'	20.000	659	13	w25/20	X	18/8 —	cone off at welding seam
L71B5	10'/21.5'	20.000	476	10	Fernox	—	18/8 —	hood off
L71B6	10'/21.5'	20.000	400	8	w25/20	X	18/8 —	cone off at welding seam
L71B6	10'/21.5'	20.000	230	5	w25/20	X	18/8 —	—
L71B6	10'/21.5'	20.000	526	11	c25/12	—	18/8 —	—
L71B7	10'/21.5'	20.000	735	15	w25/20	—	18/8 —	—
L71B7	10'/21.5'	20.000	173	3	c25/12	—	18/8 —	—
L71B7	10'/21.5'	20.000	822	16	c25/12	—	iron —	—



01 - 523 A4 - 1×1 mm

ESSELTE 4446

Day	Accum. hours	Heat supply: air mm	propane mm	BTU/h	exhaust gas volume std ft³/min	Exhaust gas circulation: jet open turns	compressor mm	cuft/min	Temp. °F: exh. gas	cone	Remarks
5/26	–	94	16	20,000	2.75	–	–	–	–	–	started
5/27	24	94	16	〃	〃	–	–	–	–	–	
5/28	48	94	16	〃	〃	–	–	–	–	–	
5/29	72	94	16	〃	〃	–	–	–	–	–	
5/30	96	94	16	〃	〃	–	–	–	–	–	
5/31	120	93	16	〃	〃	–	–	–	–	–	
	122	93	16	〃	〃	–	110	100% 2.2	240	–	circ. st.
	124	92	16	〃	〃	–	–	–	240	–	circ. shut
	130	91	16	〃	〃	1/4	–	–	230	825	jet op.
	136	91	16	〃	〃	1/2 (or 1/4?)	–	–	241	975	
6/1	144	92	16	〃	〃	1	–	–	231	780	
	149	91	16	〃	〃	1 1/4	–	–	225	820	burner out jet closed
	150	91	16	〃	〃	–	50	1.03	238	815	circ. st.
	152	91	16	〃	〃	–	70	1.46	243	810	circ. incr.
6/2	164	92	15	〃	〃	–	52	1.07	258	765	
	168	91	16	〃	〃	–	80	1.66	240	776	
	170	91	15	〃	〃	–	80	1.66	242	810	Burner out at 100 mm
	172	93	16	〃	〃	–	40	0.80	238	840	
	174	93	16	〃	〃	1/4	45	0.92	232	835	Jet opened
	178	94	16	〃	〃	1/2	60	1.26	231	900	
	180	94	16	〃	〃	3/4	60	1.26	238	920	
	183	95	16	〃	〃	1	60	1.26	224	775	burner out at 1 turn relit.
6/3	193	93	16	〃	〃	3/4	–	–	212	720	
	197	92	16	〃	〃	3/4	20	0.30	266	820	
6/4	211	92	16	〃	〃	3/4	20	0.30	220	865	

25,000 〃 〃 ~ 2.70 〃

33,000 〃 〃 ~ 3.60 〃

day	hours	air mm	prop. mm	BTU/h	[illegible]	jet op. turns	[illegible]	cu ft/h	exh. gas	cone	
6/4	212	93	16	20,000	2.15	3/4	58	1.20	223	860	
	215	93	16	20,000	2.15	3/4	58	1.20	220	858	
	217	126	19	25,000	2.70	1/4	—	—	256	965	higher input jet opened
	220	124	18	"	"	1/2	—	—	248	905	
	223	124	18	"	"	3/4	—	—	250	905	
	228	124	18	"	"	1	—	—	246	880	Jet opened 1 turn, Burner out, Reset at 3/4
6/5	238	126	17	"	"	3/4	—	—	243	889	
	239	126	17	"	"	—	20	0.30	258	968	
	242	124	18	"	"	—	40	0.80	267	984	
	245	125	18	"	"	—	60	1.26	269	955	
	252	131	19	"	"	3/4	—	—	251	926	
6/6	258	129	18	"	"	—	20	0.30	261	950	
	261	130	18	"	"	—	40	0.80	263	961	
	262	130	18	"	"	—	80	1.66	260	954	
	266	122	18	"	"	—	80	1.66	268	915	
	268	130	18	"	"	3/4	—	—	253	967	
	271	130	18	"	"	—	100	2.08	280	900	
	273	131	18	"	"	—	100	2.08	128	430	Burner probably out.
6/7	275	131	18	"	"	3/4	—	—	217	330	Burner off
	Burner pulled, checked and reset.										
	286	165	20	33,000	3.60	—	20	0.30	238	770	Burner started
	289	165	20	"	"	—	40	0.80	267	680	
	292	162	20	"	"	—	60	1.26	298	870	
	295	160	20	"	"	1/4	~~60~~	—	312	910	Circ. off. Jet opened.
	304	157	20	"	"	—	100	2.08	268	965	Circ. on. jet closed.
	307	156	20	"	"	3/4	—	—	306	904	Circ. off. jet op.

Test finished.

25,000 " " ~ 2.70 "

33,000 " " ~ 3.60 "

Exhaust gas

20 mm	= 15	CFH PROPANE 30 PSIA	= 0.30 scuft/min
40	40		0.80
60	63		1.26
80	83		1.66
100	104		2.08

Propane

16 mm	= 7.5 - 8 CFH PROPANE
18 mm	= 10
20 mm	= 13

Air

96 mm	= 2.06 SCFM air
116 mm	= 5.00

$V_1 = V_0 \cdot \sqrt{\frac{\gamma_0}{\gamma_1}}$; γ propane, 30 psia $= c \cdot 44 \cdot 30$

γ exhaust gas, 30 psia $= c \cdot \left(\frac{20}{100} \cdot 44 + \frac{80}{100} \cdot [illegible]\right) 30 = c \cdot 31.2 \cdot 30$

$\therefore V_1 = V_0 \cdot \sqrt{\frac{44}{31}} = \underline{V_0 \cdot 1.2}$,

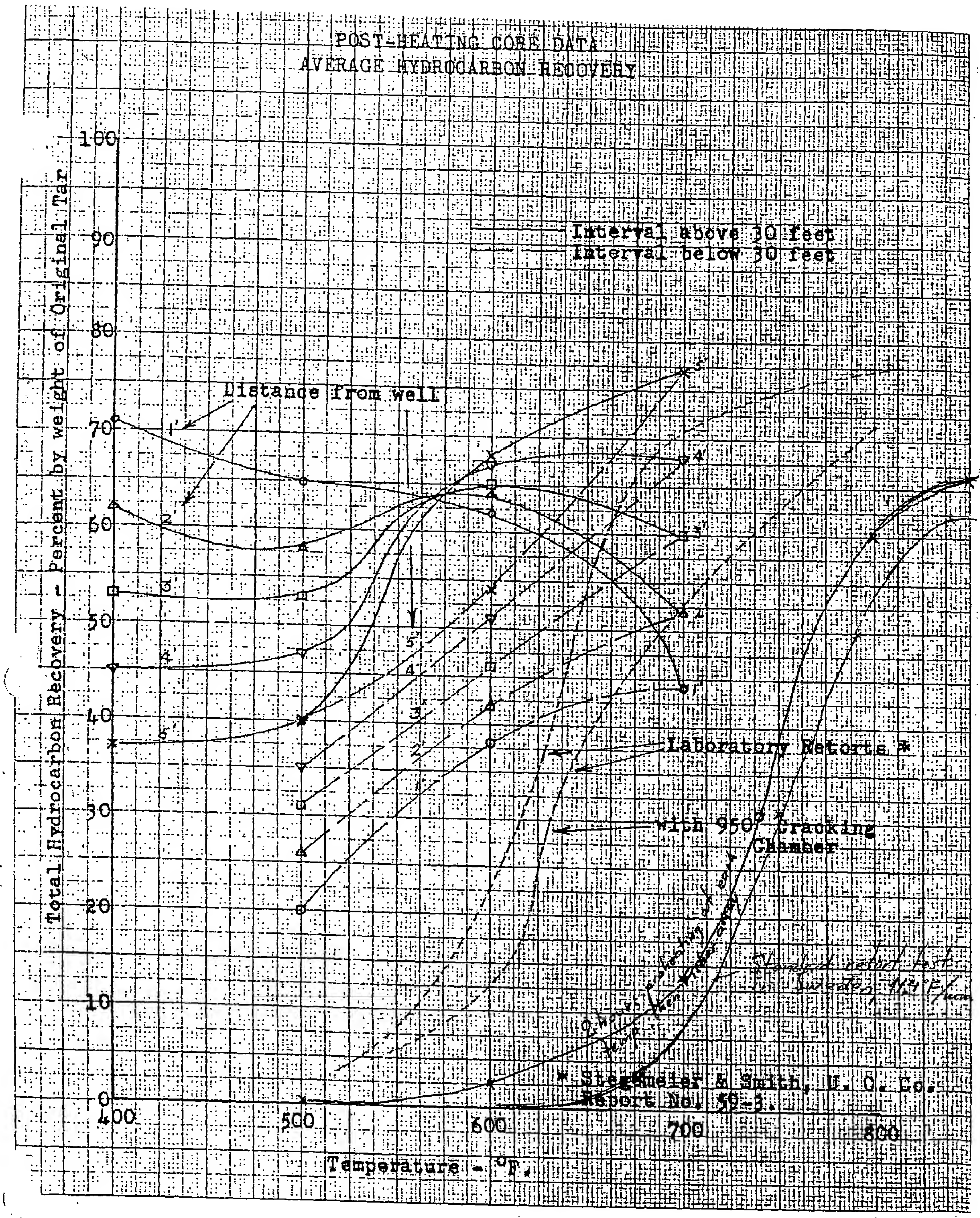
$\therefore$ 20 mm exhaust gas = 15 cuft/hour propane $= 15 \cdot 1.2 \cdot \frac{1}{60} = \underline{\underline{0.3}}$ cuft exh/min

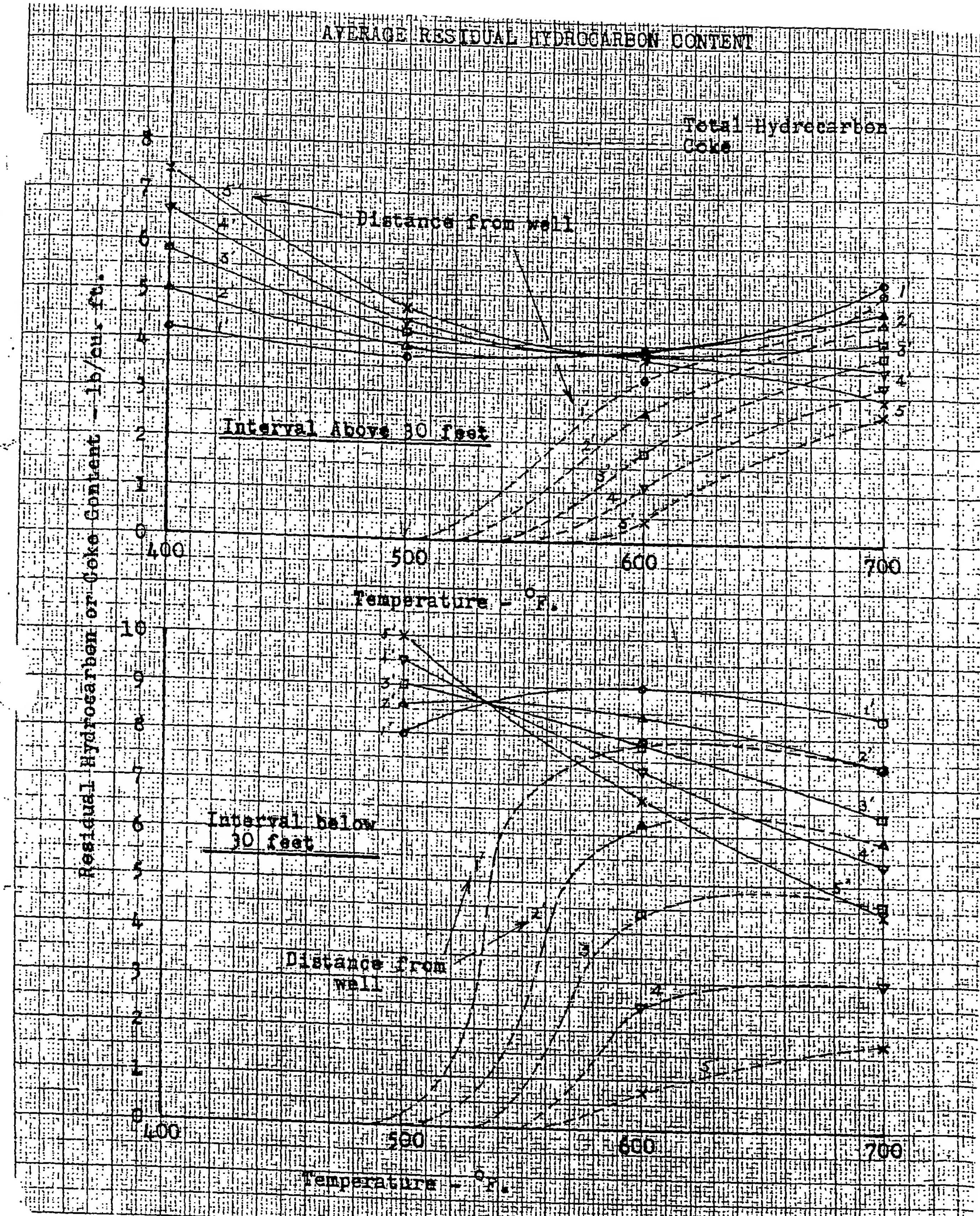
Total amount of exhaust gas.

at 20,000 BTU/h : ~ 2.15 stcuft/min

25,000 " : ~ 2.70 "

33,000 " : ~ 3.60 "


POST-HEATING CORE DATA
AVERAGE HYDROCARBON RECOVERY
Interval above 30 feet
Interval below 30 feet
Distance from well
1'
2'
3'
4'
5'
Laboratory Retorts *
with 950° Cracking Chamber
* Stegemeier & Smith, U. O. Co. Report No. 59-3.
Total Hydrocarbon Recovery - Percent by weight of Original Tar
Temperature - °F.
0
10
20
30
40
50
60
70
80
90
100
400
500
600
700
800



AVERAGE RESIDUAL HYDROCARBON CONTENT
Total Hydrocarbon
Coke
Distance from well
Interval Above 30 feet
Residual Hydrocarbon or Coke Content — lb/cu. ft.
Temperature - °F.
Interval below 30 feet
Distance from well
Temperature - °F.

It is assumed that the specific heat of tar sand is additively composed of the specific heats for each of the components, quartz sand, tar and water. These are:

(Perry: Chem. Engineer's Handbook, 3rd edition, p. 223)

Quartz: true specific heat at $t°C$ (valid between 0 and 575°C) =

$$S = \frac{1}{60.06} \cdot \left[10.87 + 0.0087\,(t+273) - \frac{241{,}200}{(t+273)^2}\right]$$

From the equation the following values were calculated:

$t°C$	0	100	200	300	400	500	(600)
true spec. heat, cal/g.°C	0.167	.206	.232	.257	.270	.286	(.302)
mean — " — betw. 0 and t°C, cal/g.°C	0.167	.189	.204	.217	.228	.237	(.247)

C. S. Cragoe (Thermal Properties of Petroleum Products, Misc. Publ. Bur. of Standards, No. 97, 1929)

Tar: according to ~~equation in Ref.~~ the true specific heat for petroleum products of a specific gravity of d (measured at 60/60°F) is $= S = \frac{1}{\sqrt{d}} \cdot [0.403 + 0.00081 \cdot t]$ cal/g.
Assuming that the formula is valid also for tar, which has a gravity of $d = 1.06$, we obtain:

$t°C$	0	100	200	300	400	500	600
true spec. heat cal/g.°C	0.392	.470	.548	.626	.704	.782	.860
mean — " — between 0 and t°C, cal/g.°C.	0.392	.431	.470	.509	.548	.587	.626

As the tar is pyrolyzed (decomposed) at temperatures above 300°C, the specific heats given for higher temperatures, are only theoretical.

~~The~~ Diagrams on p.   show the deviation of ~~the~~ the true heat consumption curve from the ~~curve~~, based on the mean specific heat (represented by the straight line).
Diagram 2 is a tendency curve, showing ~~#~~ how a varia

consideration ~~need~~ has to be given to the volatile pyrolysis products as they leave the formation immediately after being formed. (The same is valid for water above 100°c). The coke (carbonaceous residue) from the tar which remains with the quartz, amounts to roughly 30 % by weight of the tar and has a specific heat of:

t °C	350	400	450	500	550	600
true spec. heat, cal/g. °C	0.331	.341	.352	.361	.372	.382

The reaction heat of pyrolysis is of the order of 50 cal/gram of tar.

Water: average specific heat between 0 and 100 °C = 1.00 cal/g, °C.

Heat of vaporization = 540 cal/g at 100°C.

For heating 100 grams tar sand consisting of 90% b.w. quartz, 8% b.w. tar and 2% b.w. water, thus is required:

from 0 to 300°C (no pyrolysis assumed):

90 g quartz :	90 · 300 · 0.217 =	5859	cal
8 g tar :	8 · 300 · 0.509 =	1222	
2 g water :	2 · 100 · 1.000 =	200	
vaporization :	2 · 540 =	1080	
	total	8361	cal

corresponding to a mean specific heat of $\frac{8361}{300 \cdot 100} = \underline{\underline{0.27}}$ cal/g °C

from 0 to 400°C (pyrolysis assumed to take place at 350°C):

90 g quartz	90 · 400 · 0.228 =	8108	cal
8 g tar :	8 · 350 · 0.529 =	1481	
pyrolysis :	8 · 50 =	400	
2.7 g coke :	2.7 · 50 · 0.331 =	45	
2 g water :	2 · 100 · 1.000 =	200	
vaporization :	2 · 540 =	1080	

corresponding to a mean specific heat of $\frac{}{400 \cdot 100}$ = ... /g.°C

from 0 to 500°C:

90 g quartz:	$90 \cdot 500 \cdot 0.237$ =	10,665 cal
8 g tar:	$8 \cdot 350 \cdot 0.529$ =	1481 "
pyrolysis:	$8 \cdot 50$ =	400 "
2.7 g coke:	$2.7 \cdot 150 \cdot 0.349$ =	141 "
2 g water evaporation:	=	1280 "
total		13,967 cal

corresponding to a mean specific heat of $\frac{13,967}{500 \cdot 100}$ = 0.279 cal/g. °C

For a tar sand, consisting of 88% b.w. quartz, 10% b.w. tar and 2% b.w. water, is obtained in the same way:

0 – 300°C: mean specific heat = 0.285 cal/g. °C
0 – 400°C: — " — 0.290 — " —
0 – 500°C: — " — 0.284 — " —

Strictly, the heat transfer calculations require the use of the true specific heats at every temperature. This would, however, make the calculations very complicated. ~~and~~ In view of the generally occurring poor homogenity of a tar sand layer it is obvious that a constant avrg value can be used. For tar sands with 8-10% tar and temperatures within the 300-500°C range this value is: $s = \sim 0.28$ cal/g. °C.

~~The~~ Diagrams 1 on p. shows the deviation of ~~the~~ the true heat consumption curve from the curve, based on the mean specific heat (represented by the straight line). Diagram 2 is a tendency curve, showing ~~#~~ how a variat

tar sand.

Note. In the Blair Report on Athabasca Tar Sands (published in Edmonton 1950) the following is said about the thermal properties of the mentioned tar sand (p. 15):

"The bitumen of the bituminous sand is a viscous, asphaltic oil, displaying considerable variation in properties. Its specific gravity at 25°/25°C ranges from 1.002 to 1.027.

– – –

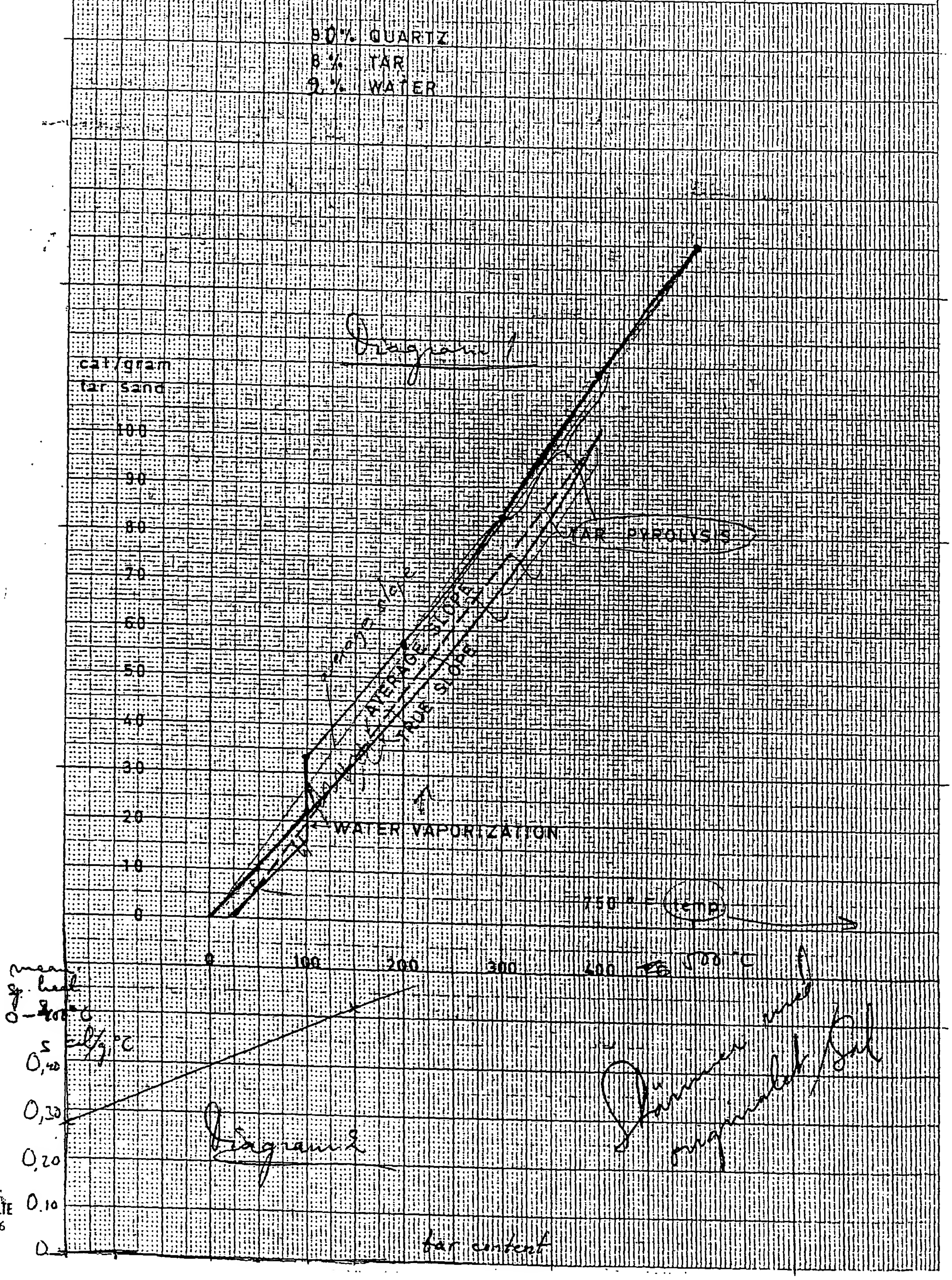
Bituminous sand in its natural state of packing weighs about 125 lb/ft^3. Its coefficient of thermal conductivity is of the order of 0.0035 in c.g.s. units. (0.0035 gm cal/sec, cm^2 (°C) = 25.8 B.t.u./hr, ft^2 (°F).) The specific heat of the mineral aggregate is 0.18 cal/gm while that of the oil is 0.35. The calorific value of the oil is 17,900 B.t.u./lb."

SSAB

TEMPERATURE OF TAR SAND FROM ~~18°~~ 0°C C TO 1° C.

AUG. 22. 1958

Sal.



91-523 A4 · 1x1 mm

ESSELTE
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10. Introduction.

The purpose of this report is to provide a basis for the design of field tests on the LINS Method for oil recovery from tar sands. At the time when these calculations were made, some of the data were known with only an unsatisfying degree of accuracy and others were completely missing. Thus a number of assumptions had to be made (described below). In this way preliminary heating patterns for the field tests have been established and the expected results have been calculated. By checking the actually obtained results against those calculated, due corrections in the used basic data can be made.

11. Pyrolysis temperatures.

When the kerogen of oil shale is heated in absence of air a decomposition (pyrolysis) starts, whereby the big molecules of the kerogen are broken down to smaller molecules (hydrogen, carbon monoxide, carbon dioxide, hydrogen sulphide, hydrocarbons from methane up, and a carbonaceous residue, which together with the inorganic parts of the shale forms a shale coke.

The decomposition temperatures depends upon the rate of heating. The lower the rate of heating, the lower is also the temperature, when the reactions start. If the shale is heated rapidly, the decomposition does not start until at higher temperatures.

Also the quantity and quality of the products are affected by the heating rate. More oil is recovered in fast than in slow heating, and a higher rate of heating results in a lower API-gravity of the oil and a higher percentage of unsaturated compounds.

perature" for three different heating rates for Swedish oil shale. It is obvious that the rate of heating constitutes a considerable difference in the pyrolysis conditions in a retort furnace and an in-situ field.

From the resemblance in behaviour in preliminary small-scale laboratory tests ~~in an in-situ field~~ between oil shale and tar sand it is assumed that the pyrolysis temperatures in an in-situ field in tar sand will be the same as for oil shale. Thus it is assumed in all the following calculations that if the tar sand temperature is:

below 600°F:	0%	of the recoverable oil has been obtained
600-650°F:	25%	— " —
650-700°F:	50%	— " —
700-750°F:	75%	— " —
above 750°F:	100%	— " —

In a field operation part of the oil and the gas spreads to the surroundings and is not recovered. The actual recovery is thus determined by field conditions (size of operation, permeability of the formation and of the surrounding rock etc.)

The quality of the recovered oil can be changed by dissolution of a smaller or larger amount of unpyrolyzed ~~tar~~ tar therein.

The tar content of the sand does probably not influence the shape of the curves, which are given in % recovered oil of all recoverable oil.

Principle. A jet is a device, whereby part of the movement energy content of one fluid is transferred to another fluid or to another part of the same fluid. In practice the first fluid is allowed to flow through a nozzle, surrounded by the fluid to be moved, whereafter the two fluids together flow through a tube, called ~~diffu~~ throat and a cone-shaped diffusor.

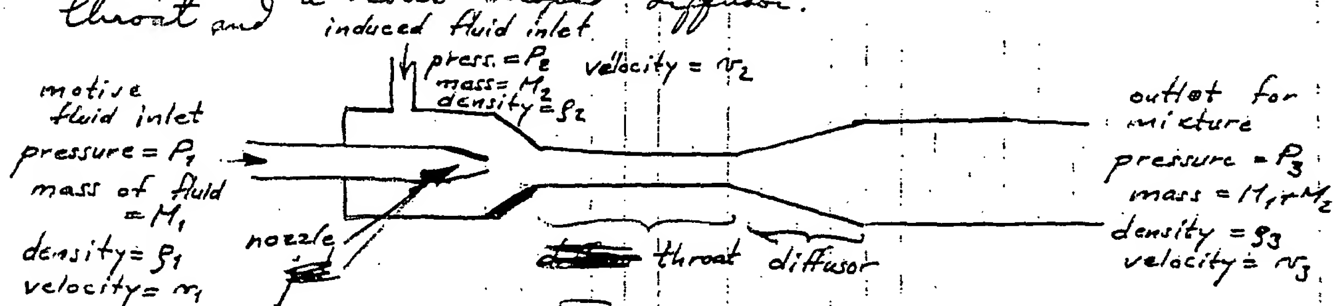


Figure 1.

The jet equation.

As the flow through the jet is adiabatic, the law of energy conservation gives (with symbols from Figure 1 above).

$$\underbrace{\frac{M_1}{\rho_1}\cdot P_1 + M_1\cdot\frac{v_1^2}{2}}_{\text{energy of fluid 1}} + \underbrace{\frac{M_2}{\rho_2}\cdot P_2 + M_2\cdot\frac{v_2^2}{2}}_{\text{energy of fluid 2}} = \underbrace{\frac{M_1+M_2}{\rho_3}\cdot P_3 + (M_1+M_2)\cdot\frac{v_3^2}{2}}_{\text{energy of outgoing mixture}} \quad (1)$$

This is the basic jet equation.

~~If the two fluids are gases the $M\cdot\frac{v^2}{2}$ - terms can almost always be as being much smaller than the $\frac{M}{\rho}\cdot P$ - termes. For instance, for air of atmospheric pressure, flowing in a pipe with a velocity of 5 ft/second, only 0.0014 % of its total energy content is due to movement energy. Thus, the equation can be written:~~

$$\cancel{\frac{M_1}{\rho_1}\cdot P_1 + \frac{M_2}{\rho_2}\cdot P_2 = \frac{M_1+M_2}{\rho_3}\cdot P_3}$$

~~pressure drop ratio $\frac{P_1-P_3}{P_3-P_2}$ is denoted P_R we obtain:~~

If the mass ratio $\frac{M_2}{M_1}$ is denoted M_R the equation can be written:

$$M_R = \frac{\frac{P_3}{\rho_3} + \frac{v_3^2}{2} - \left(\frac{P_1}{\rho_1} + \frac{v_1^2}{2}\right)}{\frac{P_2}{\rho_2} + \frac{v_2^2}{2} - \left(\frac{P_3}{\rho_3} + \frac{v_3^2}{2}\right)} \qquad (2)$$

Under certain simplifying limitations M_R can be ~~found~~ expressed as a function of the pressure drop ratio $\frac{P_1-P_3}{P_3-P_2} = P_R$ only. Empirically found curves, obtained by ______ on air/air- and air/steam-jets are reported in ______ from which article the charts ~~on next page~~ page 2 and 3 was obtained. Design rules, recommended by the same authors, are shown on page .

Testing of a flue gas jet. With no other empirical test data than those of ______ available, his chart and design rules are used for the construction of a test jet for flue gas recirculation. The inlet tube and the jet throat were made of standard 1/4" pipe and the aperture of the nozzle was 0.180 inches. The inlet chamber for the induced gas had four circular openings, each with an area of 0.0125 sq. in. or together 0.050 sq. in. The nozzle could be screwed into the chamber, so that the opening between the nozzle and the end of the throat could be varied, thereby varying the amount of induced gas. Compare drawing on page 5. Turning the nozzle 360° (one thread) moved the nozzle 0.05 inches.

Burner L5

Burner diam. = 1".

Supplied BTU:s = 42.000 BTU/h.

Amount of propane: 17.2 stcuft/h

——"— air: 410 —"—

——"— exhaust gases: 440 —"—

Amount of exhaust gases to be recirculated: assume 30% = = ~150 stcuft/h.

$$M_R = \frac{M_S}{M_J} = \frac{150 \text{ stcuft exhaust gases}}{17.2 \text{ propane} + 410 \text{ air}} = $$

$$= \underline{\underline{0.30}}.$$

For this mass ratio a pressure ratio of 8–10 is sufficient.

Assume $P_R = 10$.

$$P_R = \frac{P_C - P_D}{P_D - P_S} = 10.$$

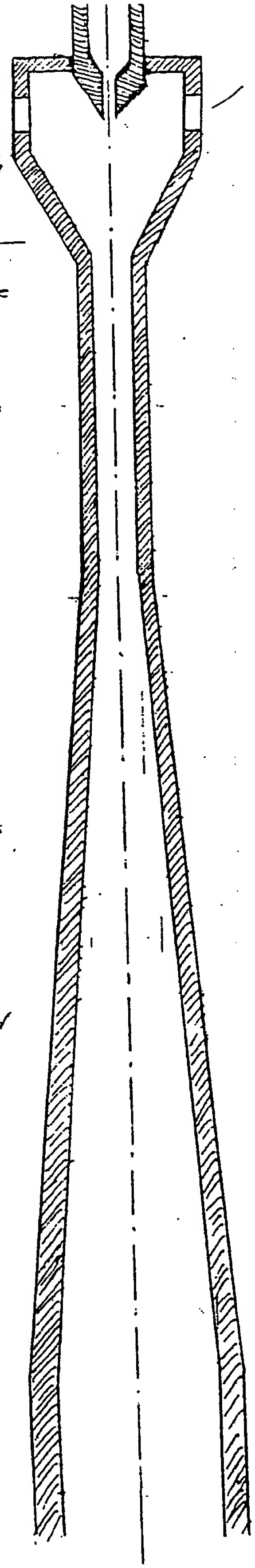
Assume $P_D - P_S = \frac{1}{2}$ psi.

Then $P_C - P_D = 5$ psi.

(Probably $P_D - P_S$ is much less than ½ psi)

For this pressure ratio the optimum diameter ratio will be $D_R = 3.–4.$

Thus the jet diameter should be about 1/16" if the diffuser throat is ¼".



[illegible]

motive fluid. As the jet was mounted in a horizontal 2¼" pipe with a bottom, the induced fluid was also air, being recirculated. By measuring the flow in the annular space between the elongated jet tube and the casing by means of ~~a nozzle~~ an orifice ring inside the casing, the degree of recirculation could be observed. The air supply was controlled by a control valve and was kept constant during the test series. Only relative measurements of the gas volumes were made. A pressure difference reading of 68 ~~millimet~~ mm aq. on the orifice ~~manometer~~ U-tube was set equal to 100 volumes of air. With different distances between jet nozzle and throat the following results were obtained (test ①):

~~distance = 0.00 U-tube reading 68 mm, corresp. to 100 volumes of air~~
~~" = 0.0125" (¼ thread) 80 108.5~~

distance, inches	U-tube reading mm aq.	corresponding air volume	% recirculation
0	68	100	0
0.0125 (¼ thread)	80	108.5	8.5
0.025 (½ ")	93	117	17
0.0375 (¾ ")	105	124	24
0.050 (1 ")	120	133	33
0.075 (1½ ")	131	139	39
0.100 (2 ")	141	144	44
0.150 (3 ")	145	146	46

Cf. diagram ~~2~~ on, page 7

It is recognized that the measuring method, used in this test ~~—~~ had some limitations and sources of error. E.g. the actual volumes of air are not exactly proportional to √(U-tube reading), as has been assumed here, as the absolute pressures at the measuring point are not taken into ... of

The same test equipment and measuring methods were used as in the above-mentioned test. A number of jets with different throat lengths and throat diameters were built and tested with the following results (throat diameters are nominal inner diameters of standard iron pipe):

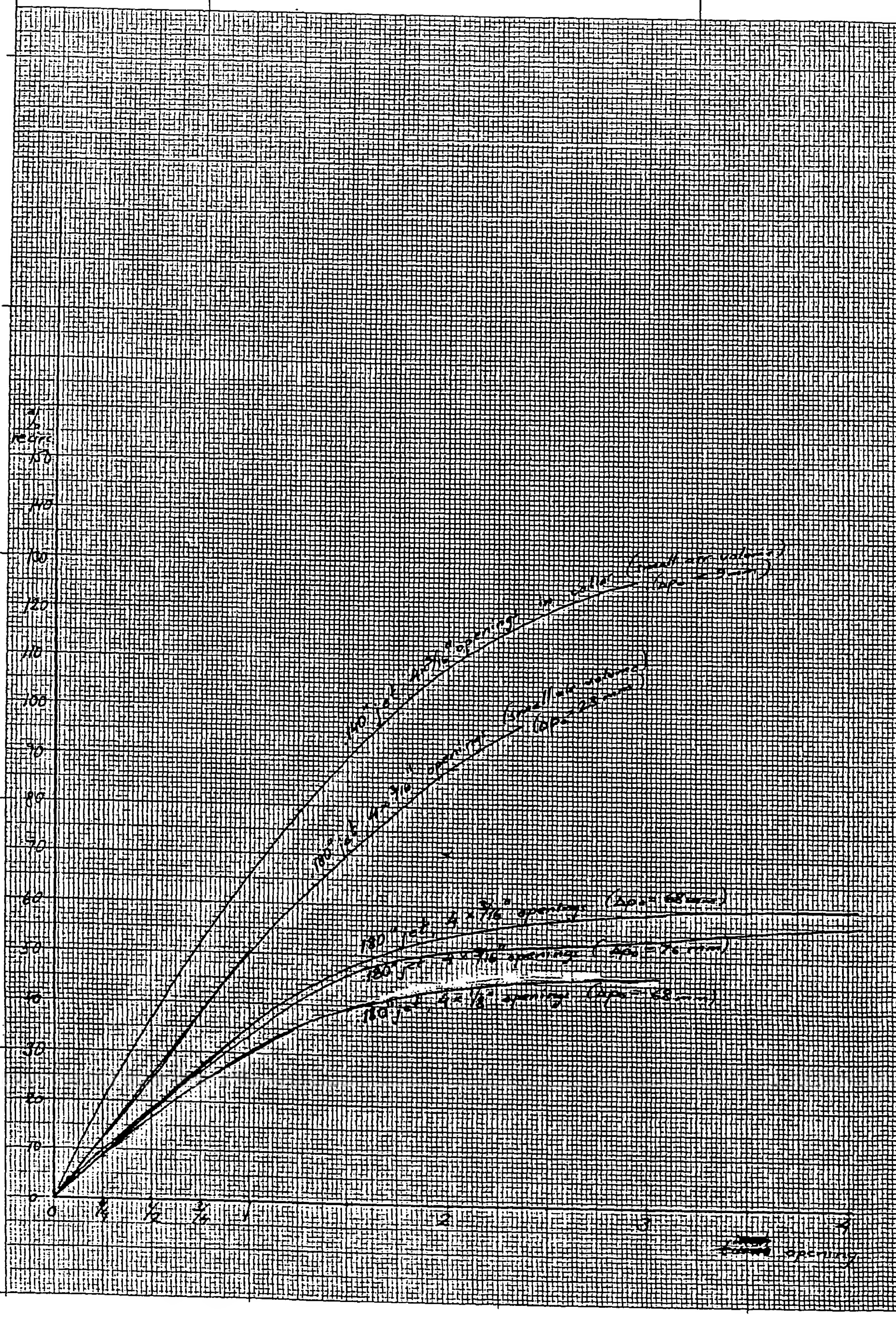
Opening in inlet	1/4" throat diameter 6" long Test run ① Δp mm	% recirc.	6" long Test run ② Δp mm	% recirc.	6" long Test run ③ (low initial velocity) Δp mm	% recirc.	54" long Test run ④ Δp mm	% recirc.
0	68	0	76	0	25	0	20	0
0.0125	82	10	92	10				
0.025	95	18	104	17	32	13	20	0
0.0375	112	28	120	26	39	25	20	0
0.050	126	36	140	36	46	36	20.5	1
0.075	146	47	160	46	54	48		
0.100	161	54	170	50	62	57		
0.150	171	57	180	54				

Opening in the throat inches	3/8" throat diameter 43" long Test run ⑤ Δp mm	% recirc.	70" long Test run ⑥ Δp mm	% recirc.	204" long Test run ⑦ Δp mm	% recirc.	1/2" throat diam. 180" long Test run ⑧ Δp mm	% recirc.
0	21	0	22	0	21	0	23	0
0.0125							26	6
0.0250	28	8	26	9			29	12
0.0375	32	15	30	17			32	18
0.0500	35	21	33	22	22.5	3	32	18
0.075	38	26	34	24	22.5	3	34	22
0.100	41	31	36	28	22.5	3	36	25
0.150	43	34	37	30	22.5	3	37	27

The results are also shown ~~on~~ Diagram 3, page 9. The [illegible] ⑤, ⑥ ~~and~~ ⑦ ~~[illegible]~~ were used for Diagram 4, ~~showing~~ that the longer the throat was, the less gas was recirculated. ~~the effect of varying throat length.~~

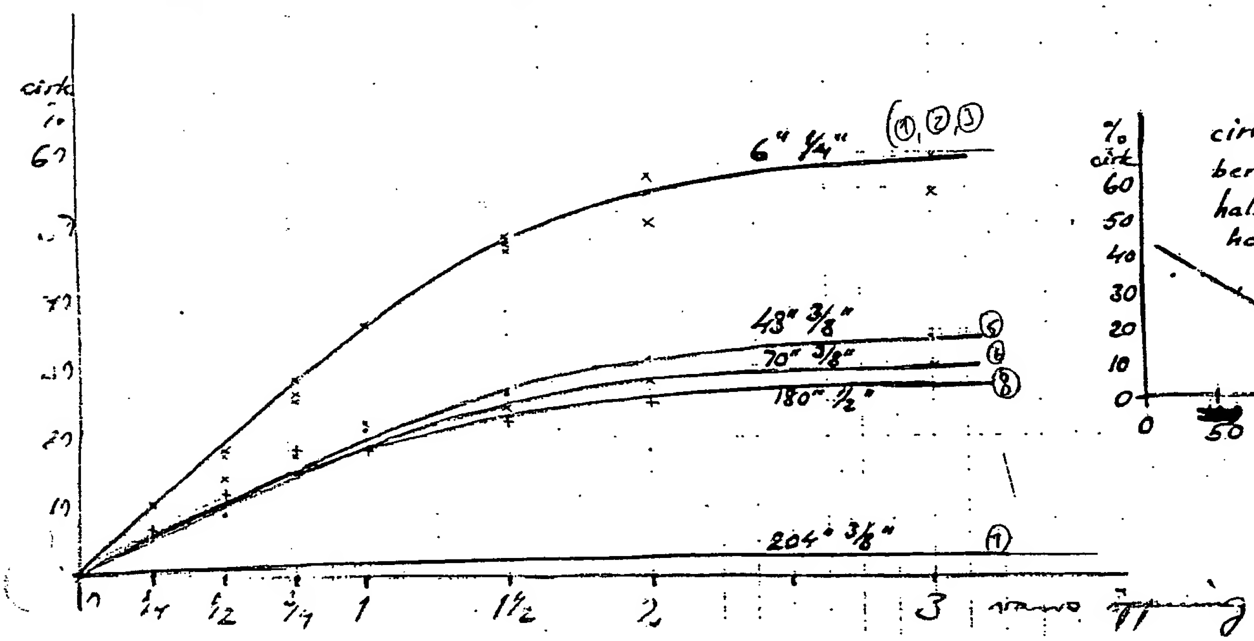
The effect of inlet openings.

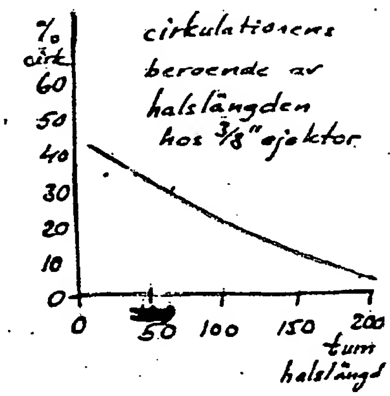
Another test series was made with 1 or 2 openings (each 0.0125 sq inch) in the inlet chamber. The gas was now natural gas which was ignited and burned at the cone, and thus the re-



514 A4
73 25 01
1 × 1 mm
ESSELTE
4441

[illegible]	Δp mm vp	% recirc	Δp	%	Δp	%	Δp	%	Δp	%	Δp	%	Δp	%	Δp	%
0	68	0	76	0	25	0	20	0	24	0	22	0	21	0	23	0
1/4	82	10	92	10											26	6
1/2	95	18	104	17	32	13	20	0	28	8	26	9			29	12
3/4	112	28	120	26	37	25	20	0	32	15	30	17			32	18
1	126	36	140	36	46	36	20,5	1	35	21	33	22	22,5	3	32	18
1 1/2	146	47	160	46	54	48			38	26	34	24	22.5	3	34	22
2	161	54	170	50	62	57			41	31	36	28	22.5	3	36	25
3	171	59	180	54					43	34	37	30	22.5	3	37	27





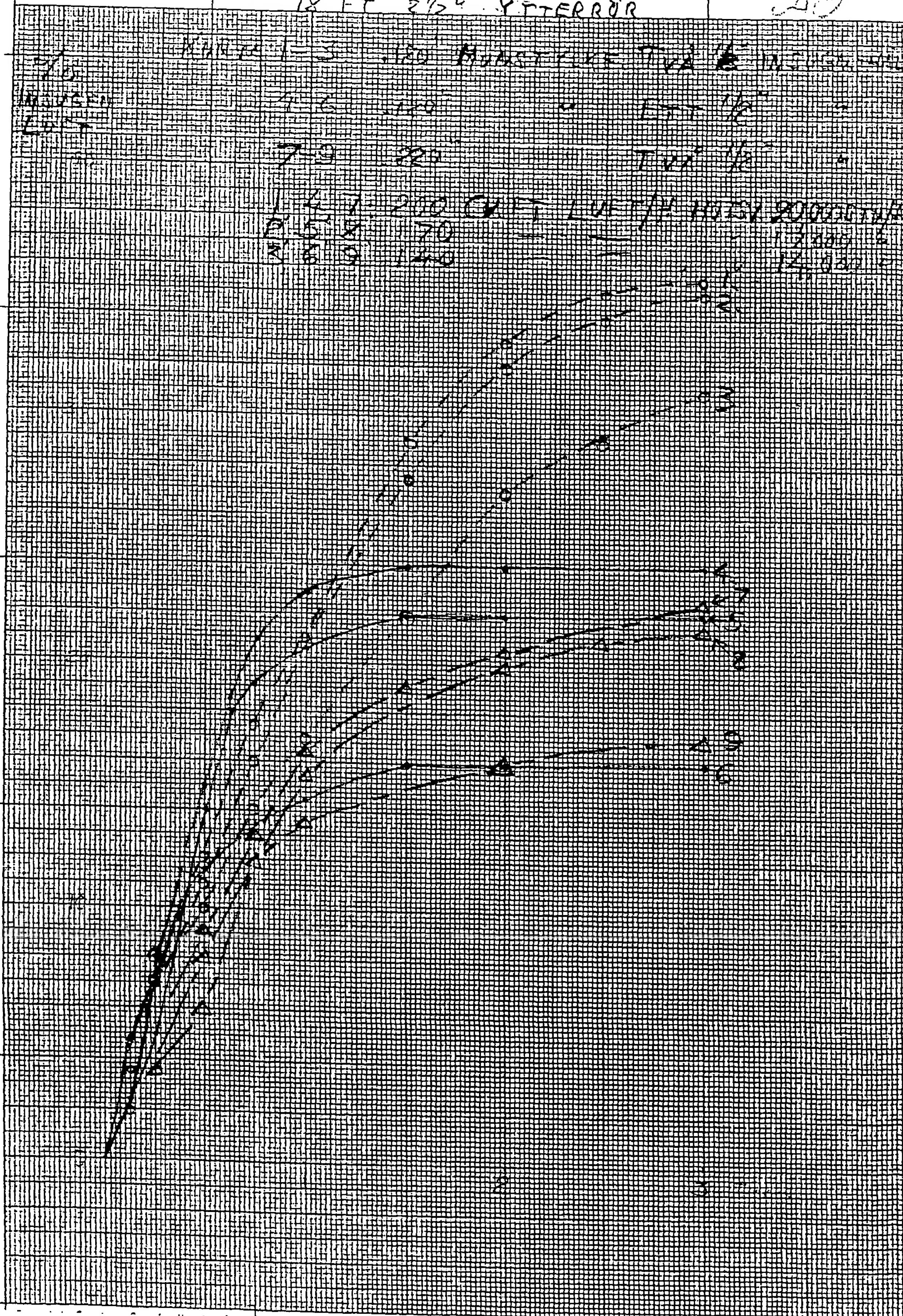
part as a series of small explosions, which made all orifice meter readings difficult and inaccurate. The following results ~~were~~ were however obtained on this 15 foot long (18 foot long) 1"-burner, inserted in an 3" casing tube.

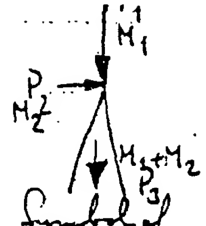
run no	nozzle aperture inch	total opening area in inlet chamber	supplied fresh ~~gas~~ ~~% of flue gas~~ gas, cuft/hr	maximum ~~recirc~~ ~~flue~~ recirculated flue gas %
9	0.180	0.0250	200	36
10	"	"	170	35
11	"	"	140	31?
12	0.180	0.0125	200	24
13	"	"	170	22
14	"	"	140	16
15	0.220	0.0250	200	23
16	"	"	170	21
17	"	"	140	17

Thus the obstructions in the flow, caused by insufficient inlet openings for the recirculated fluid, caused a reduction in the amount recirculated.

It can also be seen from the test runs 9 - 17 that the degree of recirculation is not constant but increases with increasing flow of motive fluid. No numerical relationships ~~have~~ can be calculated from the small number of tests, run so far.

1,38" STRYPFLÄNS 15 FT 1" BRANDRÖR
18 FT 2½" STIGERRÖR





Symbol of jet

Problem: If, in a [illegible] jet, [illegible] the motive fluid is increased, f.i. doubled, what happens to the amount of induced fluid?

or:

Can a constant mass ratio $\frac{M_1}{M_2}$ be maintained over the whole working range (from $M_1 = 0$ to $M_1 = \text{max.}$) of the jet?

Case 1: Jet working against atmosphere, $P_3 = 0$ atm. gauge. (All pressures measured as gauge pressures.)

If P_1 is doubled, M_1 is increased $\sqrt{2}$ times (according to the general flow equation, $V = \text{const.} \cdot \sqrt{\Delta p}$). If P_2 is kept constant then P_R is doubled and M_R is increased between 3 and 4 ~~times~~ units (e.g. from 4 to 7–8). Thus $\frac{M_2'}{M_1'} = \frac{M_2'}{1.42 \cdot M_1}$; If M_R earlier was $= 5$, it will afterwards be about 9 and we get: $\frac{M_2'}{1.42 M_1} = 9$, or $M_2' = 9 \cdot 1.42 \cdot M_1 = 12.8 \cdot M_1 = 12.8 \cdot \frac{M_2}{5} = 2.56\, M_2$. Raising M_1 $\sqrt{2}$ times thus raises M_2 more than $\cancel{10}$ 2.5 times. No constancy.

If instead M_2 is kept constant we obtain $\frac{M_2}{1.42 \cdot M_1} = 0.7 \cdot \frac{M_2}{M_1}$; If $\frac{M_2}{M_1}$ from the beginning was ~ 5, the new M_R will be $= 3.5$, which means that P_R changes from $\cancel{50}$ 130 to 66 and if M_R was 10, it changes to 7, corresponding to a P_R change from 560 to 230. P_R thus very roughly is doubled, ~~and~~ and P_2 will remain unchanged.

Case 2 Jet working with suction at atmospheric pressure. $P_2 = 0$ atm. gauge. Then P_1 is doubled and M_1 increased 1.42 times. $\left(\frac{M_2}{M_1}\right) = f\left(\frac{P_3}{P_2}\right)$; ~~If M_2 is constant, M_R is cut in half and reduced to 0.7 of its former value.~~ As P_1 is much higher than P_3, the difference $P_1 - P_3$ will be about doubled and $P_R' = \frac{P_1' - P_3'}{P_3' - P_2}$

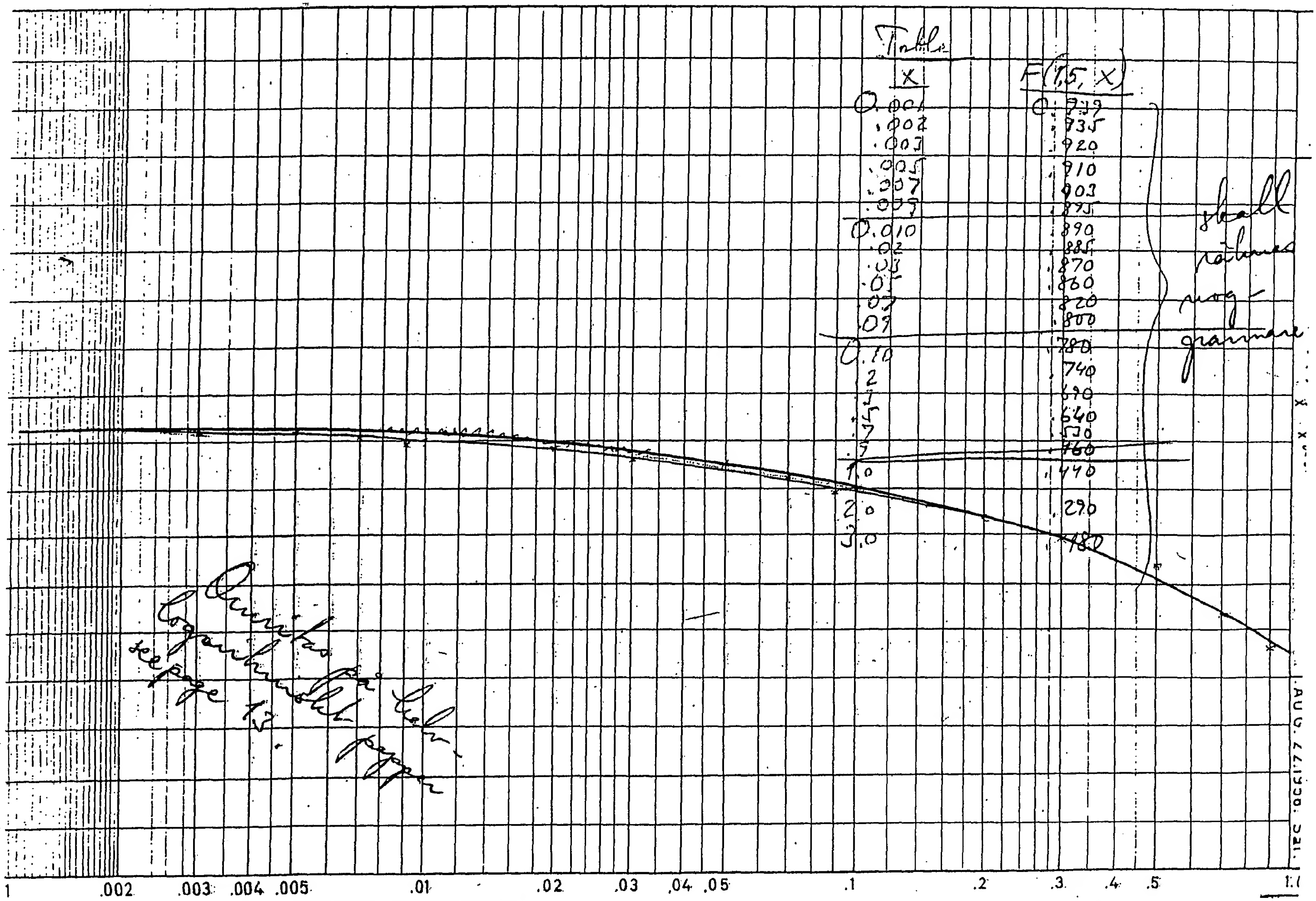
[illegible] $P' = 2(P_1 - P_3)$ [illegible]

... times are ... ...

times. Then P_R = constant and M_R also = constant and then $M_2' = 1.42 \cdot M_2$ and $M_1' = 1.42 \cdot M_1$. Proportionality.

It can be shown that an ejector always works proportional that is a certain ejector has a constant M_R and a constant P_R, independent of the amount of supplied F fluid.

(1.5, x)



Table

x	F(1.5, x)
0.001	0.939
.002	.935
.003	.920
.005	.910
.007	.903
.009	.895
0.010	.890
.02	.885
.03	.870
.05	.860
.07	.820
.09	.800
0.10	.780
2	.740
3	.690
5	.640
7	.570
.9	.460
1.0	.440
2.0	.290
3.0	.180

shall rebalance prog-gramme

Overwrite for logarithmic paper
see page 13.

Temperatures in [illegible]

Effective burner hours	Distance from burner 10"	20"	40"	60"	Effective burner hours	Distance from burner 10"	20"	40"	60"	Effective burner hours	Distance from burner 10"	20"	40"	60"
35 ft from surface					55 ft from surface					70 ft from surface				
50	110	76	63	62	62	126	98	68	62	71	124	89	69	64
334	144	114	82	70	346	156	120	95	80	355	146	126	92	77
983	154	128	98	82	995	184	166	141	108	1004	183	158	114	102
1242	155	126	100	88	1254	218	207	160	120	1263	216	192	126	113
~~1977~~	~~170~~	~~138~~	~~110~~	~~95~~	~~1789~~	~~186~~	~~172~~	~~153~~	~~125~~	~~1998~~	~~240~~	~~220~~	~~144~~	~~126~~
~~2500~~	~~165~~	~~141~~	~~114~~	~~99~~	~~2512~~	~~192~~	~~178~~	~~162~~	~~131~~	~~2521~~	~~293~~	~~230~~	~~160~~	~~140~~
~~3363~~	~~162~~	~~139~~	~~118~~	~~102~~	~~3375~~	~~192~~	~~177~~	~~162~~	~~138~~	~~3384~~	~~260~~	~~222~~	~~158~~	~~148~~
40 ft from surface					60 ft from surface					75 ft from surface				
53	114	81	64	62	65	155	116	76	64	74	118	98	70	62
337	145	116	84	70	349	(304)	213	120	85	358	140	125	88	75
986	157	130	100	83	998	378	292	168	121	1007	152	140	102	92
1245	160	132	104	89	1257	538	400	196	140	1266	164	152	110	10[illegible]
~~1980~~	~~172~~	~~142~~	~~114~~	~~98~~	~~1992~~	~~367~~	~~286~~	~~190~~	~~142~~	~~2001~~	~~180~~	~~168~~	~~124~~	~~113~~
~~2503~~	~~164~~	~~144~~	~~118~~	~~100~~	~~2515~~	~~?~~	~~325~~	~~186~~	~~152~~	~~2524~~	~~192~~	~~177~~	~~(122)~~	~~122~~
~~3366~~	~~163~~	~~144~~	~~123~~	~~103~~	~~3378~~	~~310~~	~~244~~	~~170~~	~~150~~	~~3387~~	~~184~~	~~170~~	~~138~~	~~12[illegible]~~

Burner started with 39,000 BTU/hour input

— " — off at 1294	1 hour
191	5
500	120
769	1
1009	2
1023	4
1137	10
1154	19
1433	5 ← From here 28,800 BTU/hour input
1460	7
1745	2
1844	5
2158	1
2468	5
3193	2
3455	44

Point	$\frac{r^2}{t}$	ΔT_1	$\Delta T_{1.5}$	$\frac{\Delta T_{1.5}}{\Delta T_1} = F$	x	$\frac{c^2}{t}$
~~74/10~~	~~$2.41 \cdot 10^{-3}$~~	~~56~~	~~57~~	~~.910~~	~~7~~	~~3.03~~
71/20	9.62	36	27.5	.764	140	15
[illegible]0	21.7	19	11	.58	580	27
~~358/10~~	~~.498~~	~~78~~	~~75.5~~	~~.968~~	~~[illegible]~~	~~22~~
358/20	1.98	63	52	.826	68	34
358/30	4.48	41	30	.732	22	5
358/40	7.97	26	18	.693	320	40
~~1007/10~~	~~0.177~~	~~91~~	~~86~~	~~.945~~	~~[illegible]~~	~~[illegible]~~
1007/20	.705	72	64	.889	16	23
1007/30	1.59	54	44	.815	80	50
1007/40	2.83	40	33.5	.837	57	20
~~1266/10~~	~~.141~~	~~103~~	~~101~~	~~.98~~	~~140~~	~~2~~
1266/20	.562	89	77	.865	33	59
1266/30	1.27	65	52	.800	96	76
[illegible]/40	2.25	48	41	.868	41	18

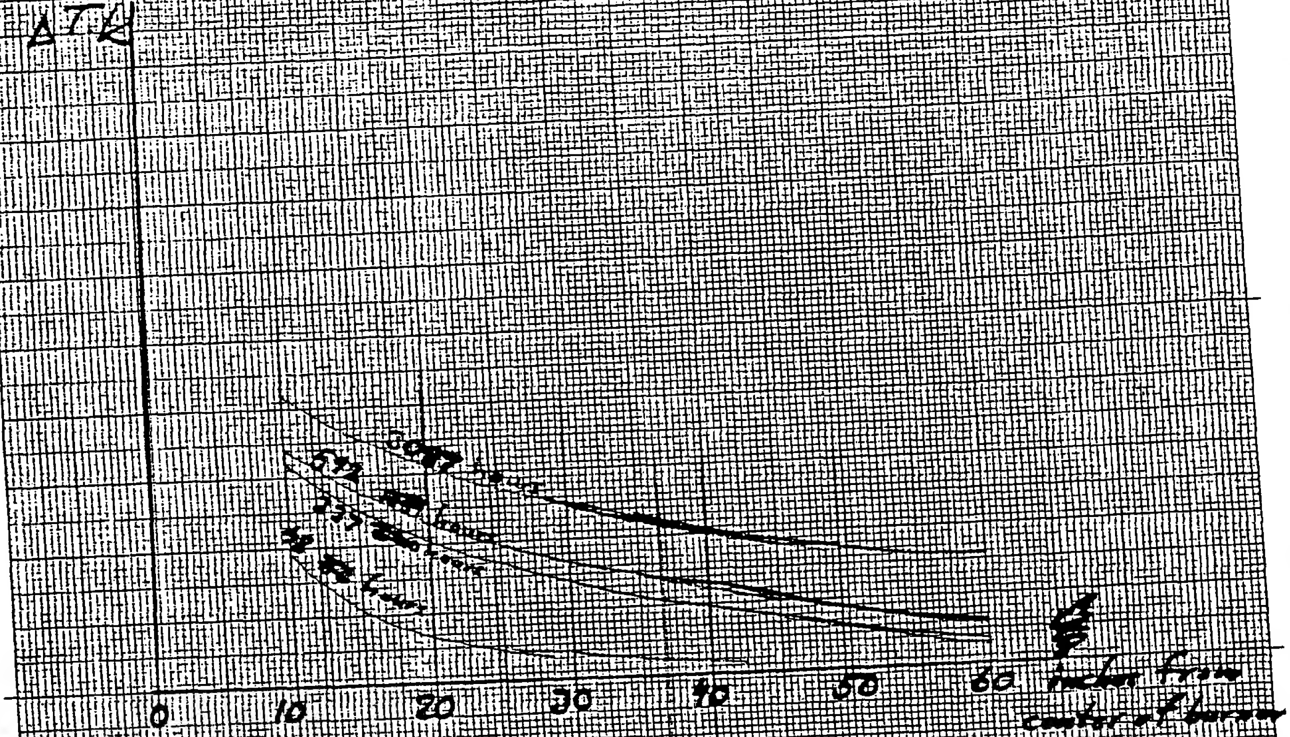
ΔT-curves ..

(core level at 55 ft)

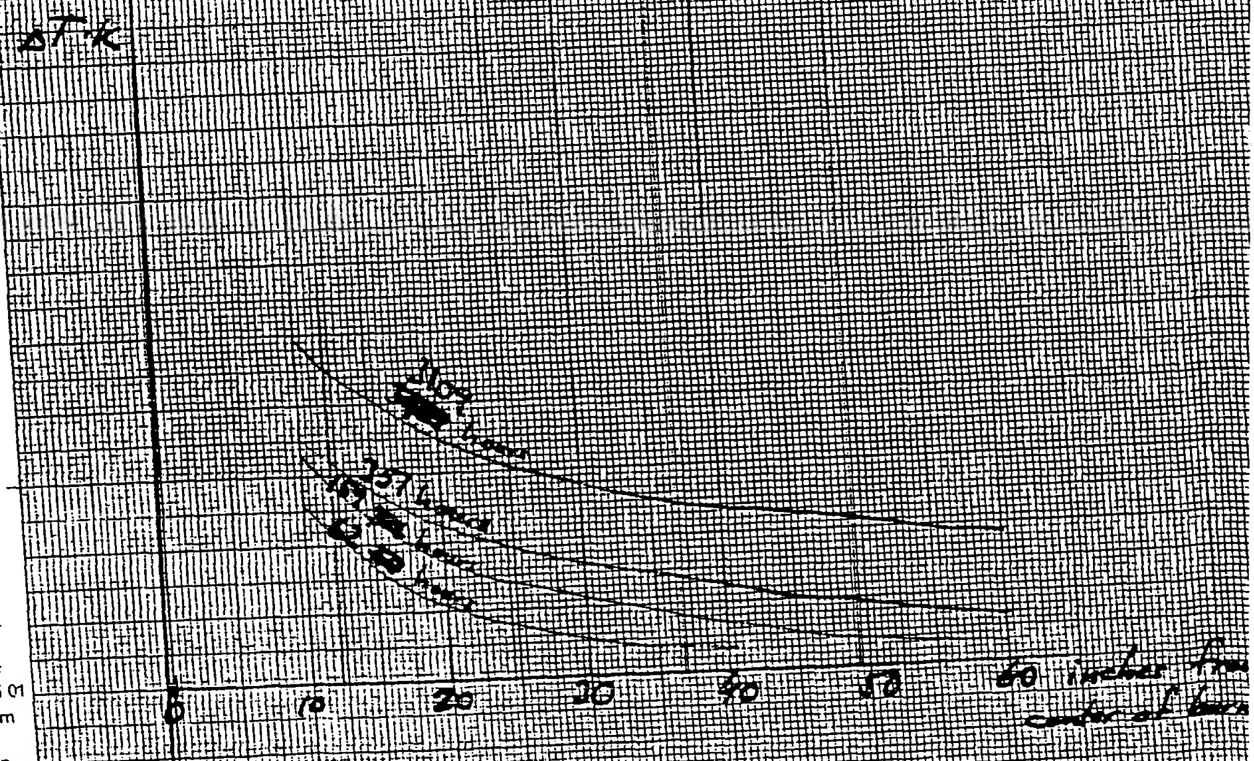
2-4

May 14 [illegible]

50 feet from surface (shale)



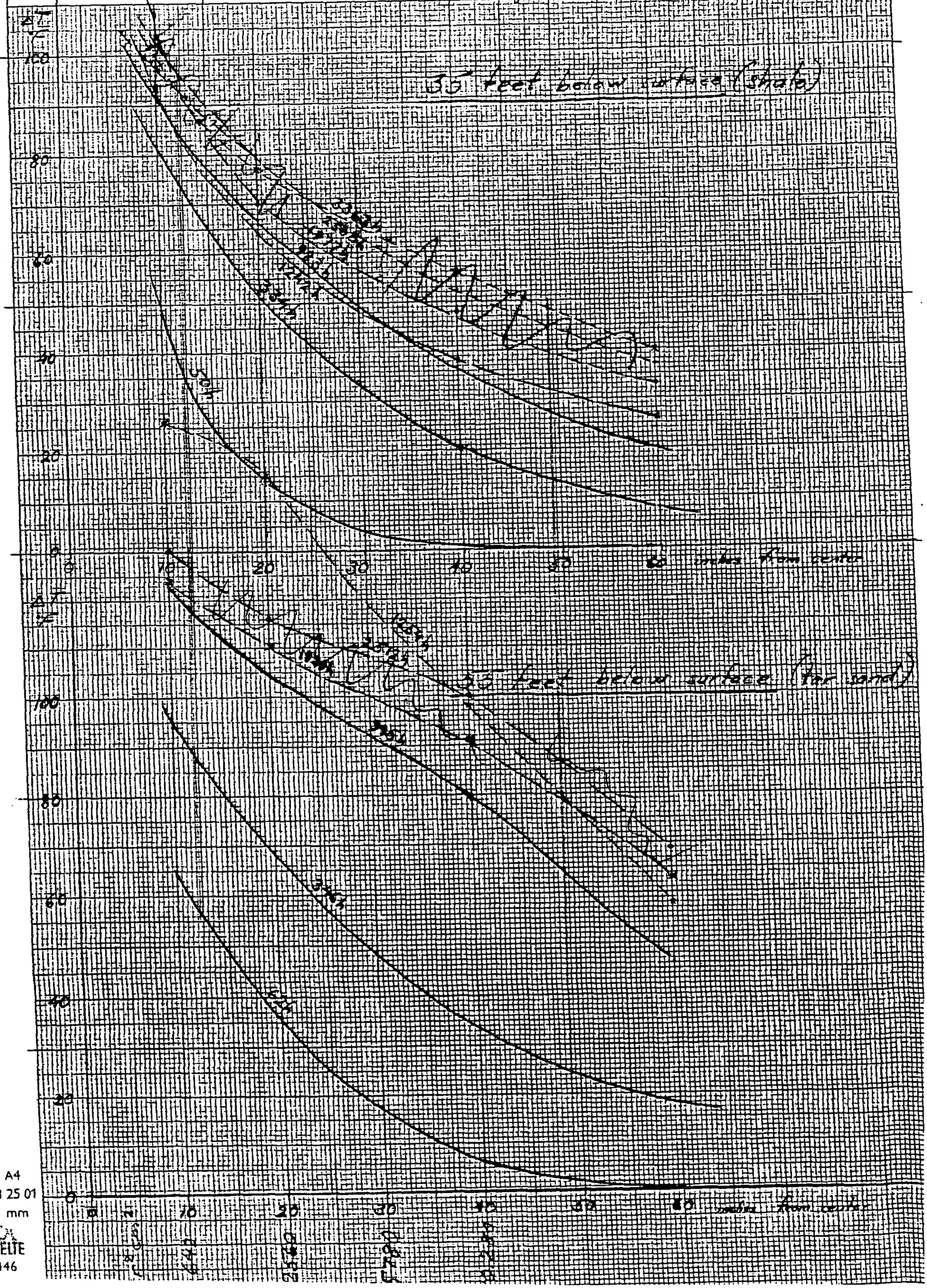
70 feet from surface (for sand)



523 A4
SIS 73 25 01
1×1 mm
ESSELTE

ΔT-curves for L4.
(cone level at 60 ft)

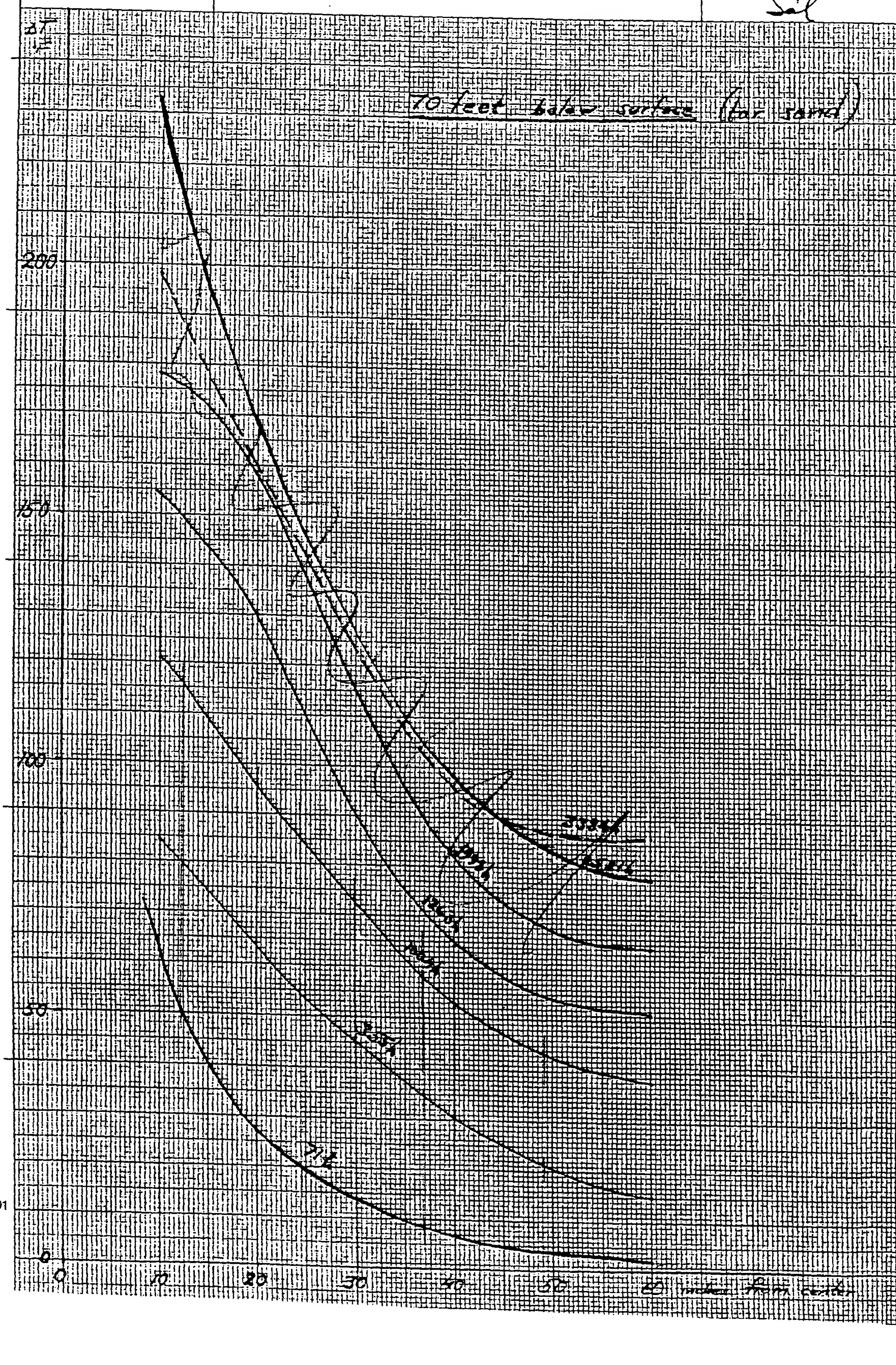
May 10, 19??
Sal

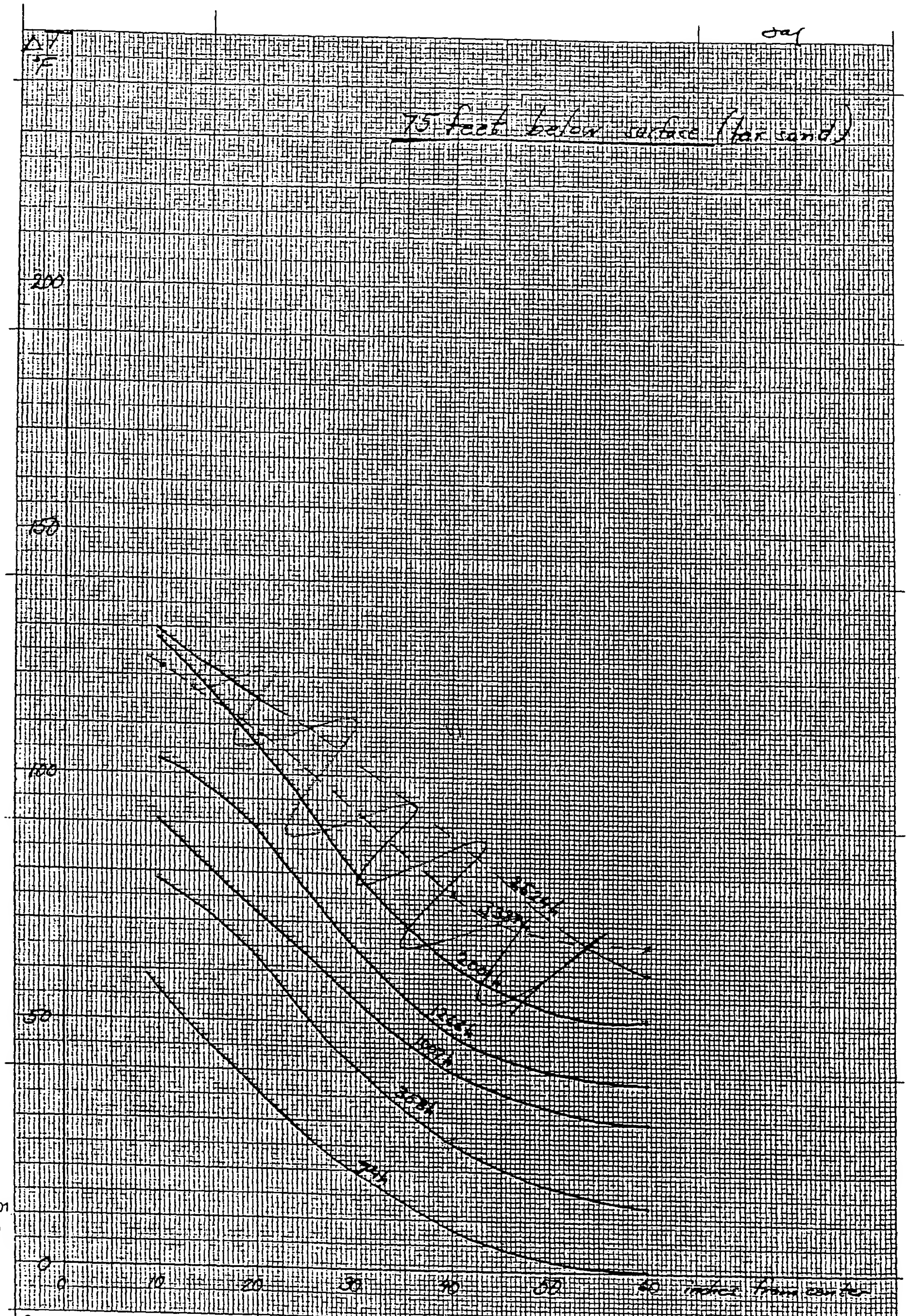


523 A4
SIS 73 25 01
1 x 1 mm
ESSELTE
4446

ΔT-curves for L4.

May 15, 1906





523 A4
73 25 01
1 mm
ESSELTE
4446

Δ	Δ^2
2,7	7,3
1,4	2,0
1,4	2,0
2,9	8,4
0,7	0,5
0,8	0,6
0,6	0,4
1,5	2,3
0,7	0,5
0,5	0,3
0,4	0,2
1,8	3,3
1,3	1,7
1,9	3,6
0,4	0,2
0	0
3,5	12,3

$\frac{21,5}{17}$ (1,3) $\frac{45,6}{17} = 2,68$

$\sqrt{2,68} = 1,64$

$\therefore \alpha = (6,1 \pm 1,6) \cdot 10^{-3}$ cm²/sec

Δ	Δ^2
11	121
20	400
29	840
75	5625
35	1225
61	3730
63	3960
72	5200
56	3150
12	144
3	9
62	3820
3	9
5	25
14	196
32	1030
43	1850
19	361
24	576
17	289
21	441
64	4100
13	169
15	225
59	3500
21	441
19	361
59	3500
3	9
34	1150
43	1850
62	3820
25	625
26	676
23	530

n=35

$\frac{54.040}{35} = 154$

$\sqrt{154} = 12,4$

$\alpha = (9,2 \pm 1,2) \cdot 10^{-3}$ cm²/sec

Omritas på halv-logaritmiskt papper

see page 13.

.5 .4 .3 .2 .1 .05 .04 .03 .02 .01 .005 .004 .003 .002 .001

10 9 8 7 6 5 4

Försök med borrbränning i tjärsand.

Allmänt.

I början av 1953 gjordes förberedande försök med brännborrning i tjärsand. Anledningen härtill var, att det visat sig svårt att få goda borrningsresultat med gängse borrmetoder vid provborrningar i tjärsanden i Albertaområdet i Canada. Tjärsanden är där mjuk och klibbig, varför den fäster på borrstängerna och kan få dessa att fastna.
Brännborrningsmetoden skulle eliminera dessa svårigheter och dessutom lämna ett koksrör efter sig som skulle förhindra borrhålet att flyta igen.

Konstruktionen av brännborret.

Brännborret består i stort sett av tre koncentriska rör, inbördes förskjutbara i sin övre del medelst packboxar och i nedre delen en brännarkrona med 8 st. dysor. Den yttre kanalen är för lufttillförsel, den mittre för gas och centrumkanalen för uppsugning av renbränd sand. Hela denna apparatur sänkes med en viss inställd hastighet medelst en utväxling driven av en elektrisk motor. Genom en arm från utväxlingsmekanismen vrider sig brännborret fram och åter 1/8 varv.

Material.

Brännborrets krona och rören 1 m närmast denna är tillverkade av eldhärdiga stål, apparaturen i övrigt i olegerat stål. Materialproblemet ligger i brännborrkronan där temp. blir hög ca. 900°C. Denna del måste säkerligen tillverkas i Kantal eller Farnox för att hålla under en längre tid. Efter ca 2 timmar har dyshålen börjat sätta igen sig på grund av flagning från godset.

Försöksresultat.

På grund av svårigheter att få hit tjärsand från Athabaska, användes samma tjärsand som tillverkades för de första förberedande pyrolysförsöken i "tjärsandsgropen". Sammansättningen var 4,5 vikts% BA-beck, 11,5 vikts% råolja från Kvarntorp 1 och 84 vikts% sjösand.
Den maximala borrhastigheten, som kunde erhållas, var ca 9 cm/h, detta vid

en propanmängd av 90 Hl/h och 2900 Hl luft/h. Hålets diameter blev ca 11 cm. Ökades sjunkhastigheten på brännaren, resulterade detta i att sanden inte hann brännas ur, och till slut stod brännaren på botten av hålet. Orsaken till den låga sjunkhastigheten torde ligga i kvartsens mycket låga värmeledningsförmåga. Sandkornen häftar gärna samman vid den höga temperaturen och kan då inte sugas upp i sugkanalen.

För att nå bättre resultat måste säkerligen sanden under själva bränningen också bearbetas mekaniskt, exempelvis med en don, fastsatt på brännarhuvudet. Det kan också tänkas, att temperaturen skall vara avsevärt mycket högre så att kvartssanden helt smältes undan och bildar en vägg av kvarts. De brännborrmetoder, som i vissa fall användes vid borrning i hårdare bergarter, måste studeras.

Kvarntorp den 25.5.1955

H.O. Tureunis.

Utl. den | av | till

FIG 1

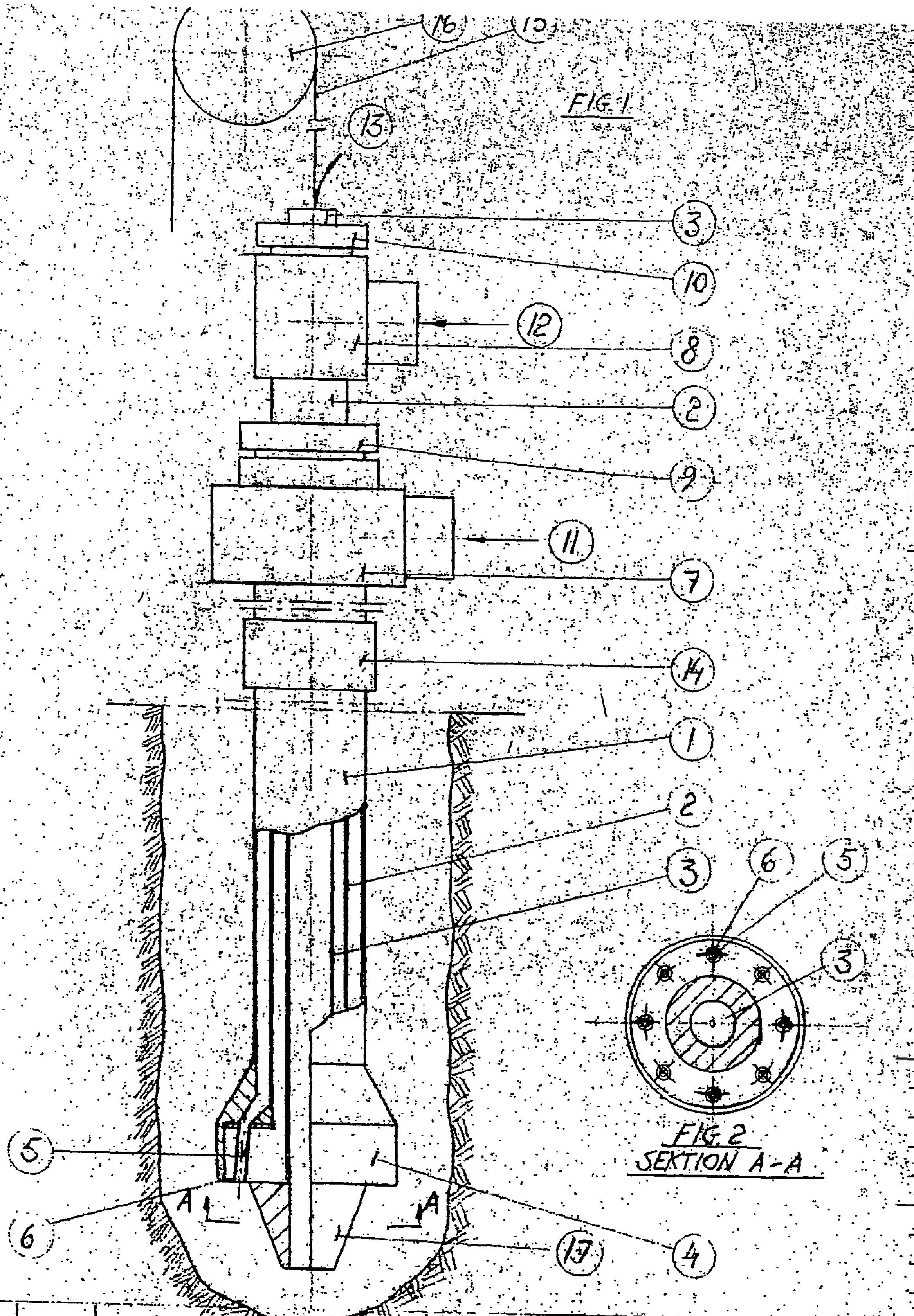


FIG 2
SEKTION A-A

B | A Antal	Detalj nr	Benämning	Beteckning Dimension	Material	Vikt	Anmärkning
Konstr.	Ritad	Kontr.	Godkänd	Stand.	Datum	Skala
Ersätter	Ersatt av					

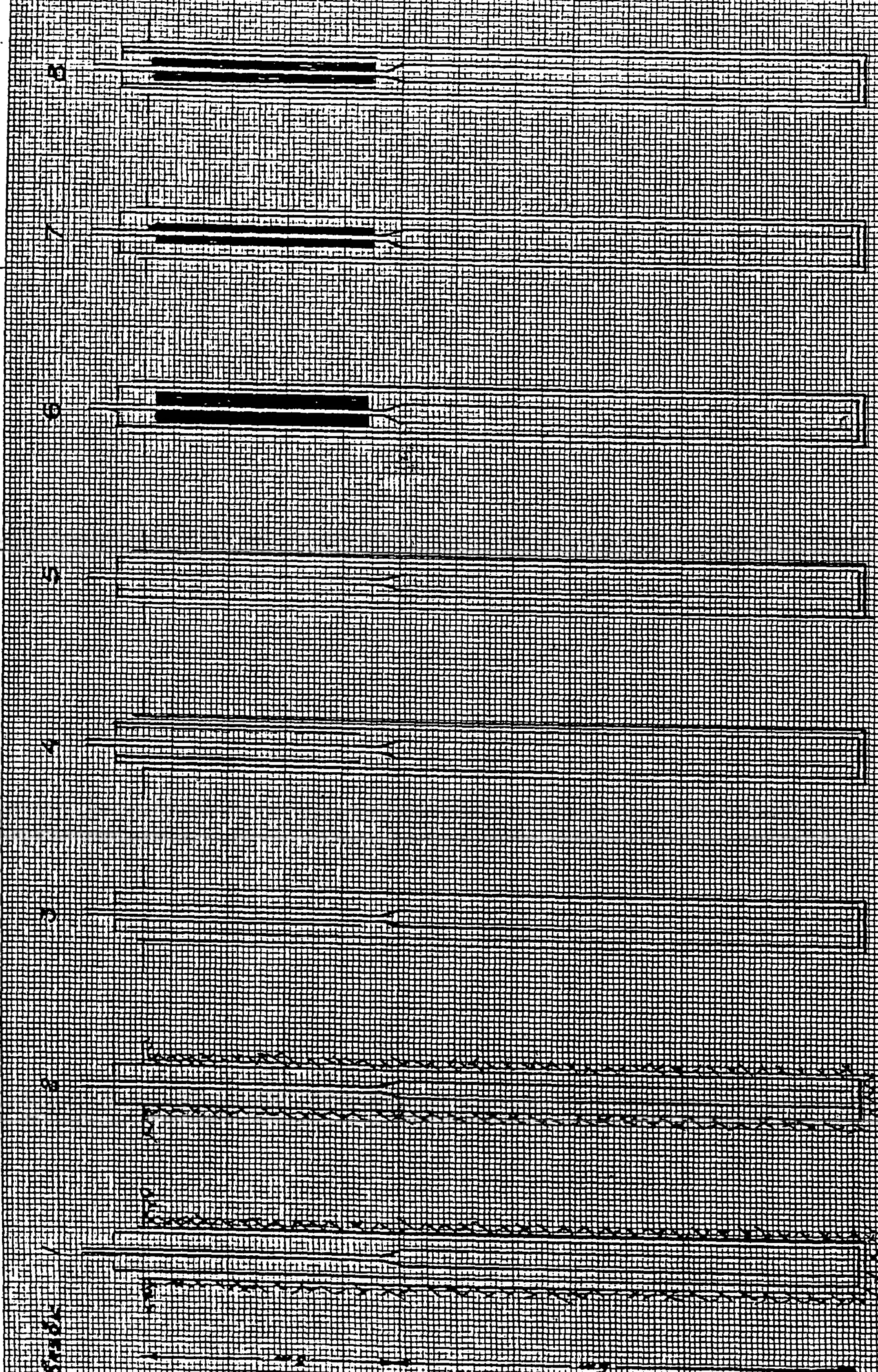
SVENSKA SKIFFEROLJE AB

C-1364

F.P.

FÖRSÖK MED VÄRMEVÄXLING.

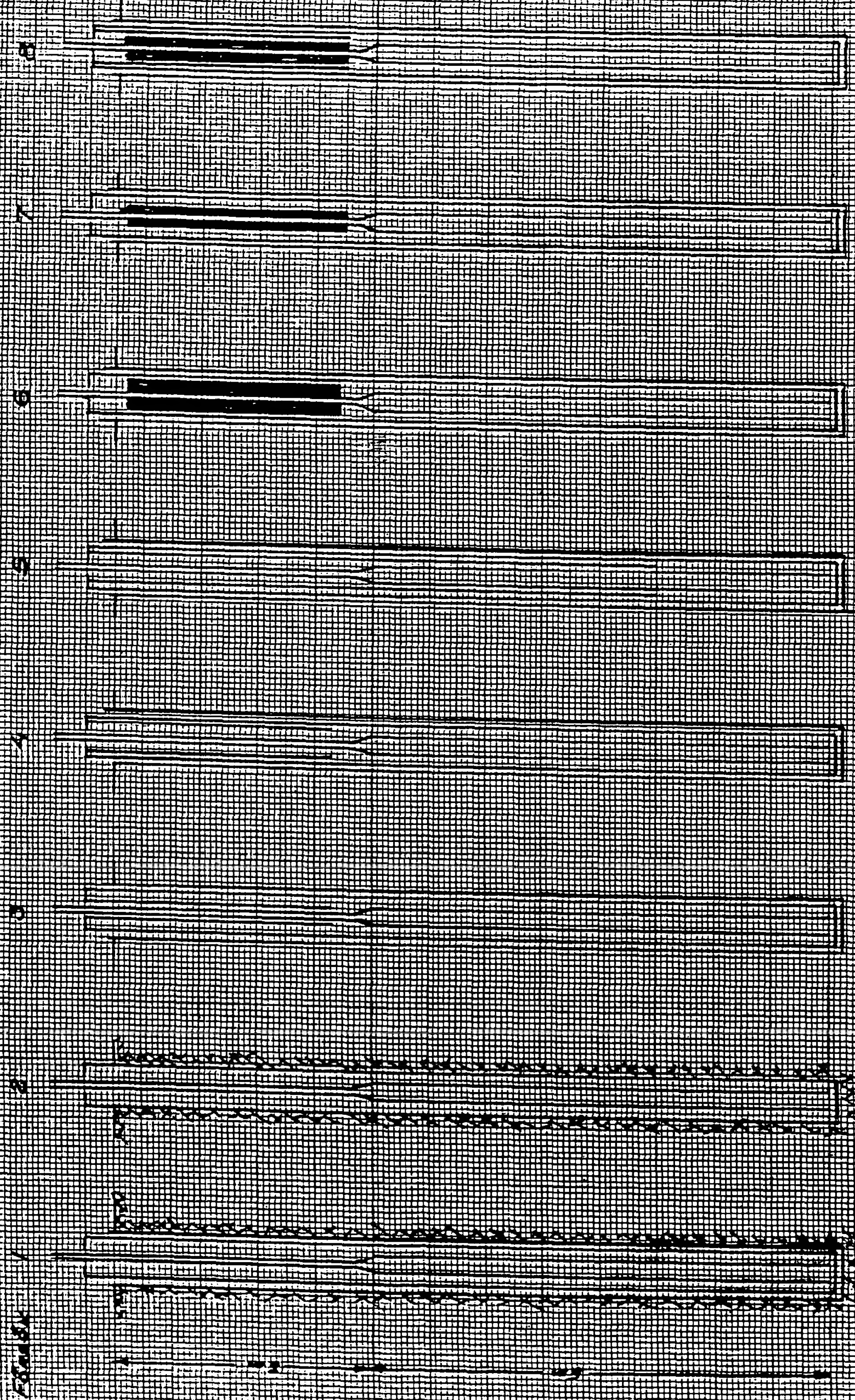
25·4.1955.



F.T.

FÖRSÖK MED VÄRMEVÄXLING.

25.4.1955.



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LIMS – brännare.
Kamflänsrör.

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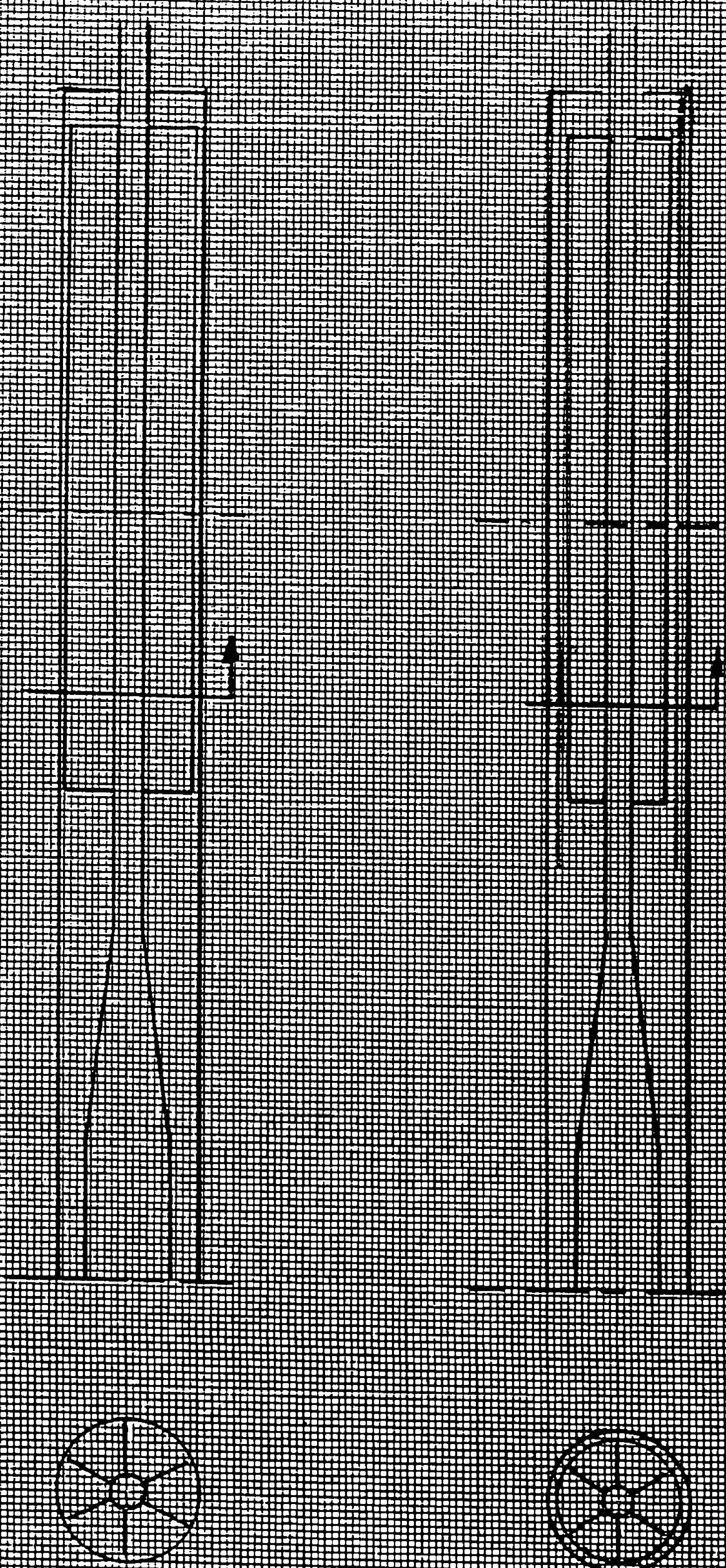


Bearb. / 194	Konstr. / 194		Reg. nr.

Heat conduction in solids.

1. Basic equation.

The heat conduction ($\frac{dQ}{d\tau}$ BTU/hour) in a solid is proportional to the ~~thermal conductivity~~ of the solid (λ BTU/ft, °F, h), the area through which the heat is conducted (A sq.ft) and the temperature gradient in the direction of flow ($\frac{dt}{dr}$; °F/ft).

Thus

$$\frac{dQ}{d\tau} = -\lambda \cdot A \cdot \frac{dt}{dr};$$

~~Heat conduct~~

Thermal conductivities for a number of solids are tabulated in table ....

2. Heat conduction when heat is stored in the conducting solid.

We consider a small volume, $dx \times dy \times dz$, in a solid body, where heat is transferred through conduction only. The temperature gradients in the three axis directions are $\frac{dt}{dx}$, $\frac{dt}{dy}$, $\frac{dt}{dz}$.

At the same time heat is stored in the volume element, causing its temperature to ~~rise~~ rise from t to $t + dt$ during $d\tau$ time units. The volume weight of the solid is ρ and its specific heat is c. Then the heat capacity of the volume element is $= dx \times dy \times dz \times c \cdot \rho$.

Further it is assumed that heat is evolved within the solid at a rate of q heat units per volume unit per time unit. During the time $d\tau$ thus is evolved $q \cdot dx \cdot dy \cdot dz \cdot d\tau$ heat units.

According to the law of constant energy we obtain:

heat, flowing into the element + heat, evolved in the element = = heat flowing out of the element + heat, stored in the element,

or

$$-\lambda_x \cdot dy \cdot dz \cdot \left(\frac{dt}{dx}\right)_x \cdot d\tau - \lambda_y \cdot dz \cdot dx \cdot \left(\frac{dt}{dy}\right)_y \cdot d\tau - \lambda_z \cdot dx \cdot dy \cdot \left(\frac{dt}{dz}\right)_z \cdot d\tau +$$
$$+ q \cdot dx \cdot dy \cdot dz \cdot d\tau =$$
$$= -\lambda_x \cdot dy \cdot dz \left(\frac{dt}{dx}\right)_{x+dx} \cdot d\tau - \lambda_y \cdot dz \cdot dx \cdot \left(\frac{dt}{dy}\right)_{y+dy} \cdot d\tau - \lambda_z \cdot dx \cdot dy \cdot \left(\frac{dt}{dz}\right)_{z+dz} \cdot d\tau +$$
$$+ dx \cdot dy \cdot dz \cdot c \cdot \rho \cdot dt$$

or, after division by $dx \cdot dy \cdot dz \cdot d\tau$:

$$-\lambda_x \cdot \frac{1}{\partial x}\left[\left(\frac{\partial t}{\partial x}\right)_x - \left(\frac{\partial t}{\partial x}\right)_{x+dx}\right] + \lambda_y \cdot \frac{1}{\partial y}\left[\left(\frac{\partial t}{\partial y}\right)_y - \left(\frac{\partial t}{\partial y}\right)_{y+dy}\right] - \lambda_z \cdot \frac{1}{\partial z}\left[\left(\frac{\partial t}{\partial z}\right)_z - \left(\frac{\partial t}{\partial z}\right)_{z+dz}\right]$$

$$= c \cdot \rho \cdot \frac{\partial t}{\partial \tau} - q.$$

or
$$\lambda_x \cdot \frac{\partial^2 t}{\partial x^2} + \lambda_y \cdot \frac{\partial^2 t}{\partial y^2} + \lambda_z \cdot \frac{\partial^2 t}{\partial z^2} = q - c \cdot \rho \cdot \frac{\partial t}{\partial \tau};$$

This is the general ~~heat~~ differential equation for heat conduction, associated with heat evolution and heat storage.

In an isotropous body the ~~heat~~ thermal conductivities are equal in all directions (and the equation can be written:

$$\frac{\partial^2 t}{\partial x^2} + \frac{\partial^2 t}{\partial y^2} + \frac{\partial^2 t}{\partial z^2} = \frac{q}{\lambda} - \frac{1}{\alpha} \frac{\partial t}{\partial \tau};$$

where $\alpha = \frac{\lambda}{c \cdot \rho}$ = the thermal diffusivity of the solid.

3. Heat conduction in an infinitely wide, plane plate.

Heat flows in only one direction and $\frac{\partial^2 t}{\partial y^2} = \frac{\partial^2 t}{\partial z^2} = 0$:

Thus:
$$\frac{\partial^2 t}{\partial x^2} = \frac{q}{\lambda} - \frac{1}{\alpha} \cdot \frac{\partial t}{\partial \tau};$$

1. Burner fuels.

The main fuel for the burners is the produced uncondensable gas from the field. Under conditions, when a make-up fuel quantity is needed, natural gas or propane can be used.

Table 20: Theoretical [illegible] components

Fuel	[illegible] value BTU/cuft	[illegible] theoret. combust. cuft/cuft gas	required: cuft gas	cuft air	total cuft	[illegible] formed: cuft CO_2	cuft H_2O (vapor)	cuft N_2	cuft total
Carbon monoxide CO	341	2.38	29.3	69.7	99.0	29.3	—	55.0	84.3
Hydrogen H_2	290	2.38	34.5	82.1	116.6	—	34.5	64.8	99.3
Methane CH_4	963	9.52	10.4	99.3	109.7	10.4	20.8	78.5	109.7
Ethane C_2H_6	1703	16.67	5.7	95.0	100.7	11.4	17.1	75.0	103.5
Ethylene C_2H_4	1631	14.29	6.1	87.1	93.2	12.2	12.2	69.0	93.2
Propane C_3H_8	2440	23.80	4.1	96.6	100.7	12.3	16.4	76.1	104.8
Propylene C_3H_6	2328	21.43	4.7	100.6	105.3	14.1	14.1	79.4	107.6

Table 21: Theoretical combustion data for three gas mixtures, corresponding to analyzed samples from field tests.

I	30% CO_2 + 2% CO + 56% H_2 + 2% C_2H_6 + 10% CH_4	292	2.62	34.2	90	124.2	15.7	27.4	70.6	113.7
II	14% CO_2 + 2% CO + 33% H_2 + 25% C_2H_6 + 26% CH_4	761	7.00	13.1	92	105.1	12.1	17.7	71.4	101.2
III	7.5% CO_2 + 55% H_2 + 7% C_2H_6 + 31% CH_4	570	5.40	17.6	95	112.6	9.1	24.1	75.0	108.2

All figures above refer to dry gases, measured at 32°F, 30 inch pressure.

Table 22: Chemical composition of fuel-air mixtures and exhaust gases.
The fuel air mixture contains the theoretical amount of air saturated with water vapour at 50°F
The exhaust gas is assumed to leave the burner ~~without~~ before any condensation of water vapour has taken place.

Gas	Average molecular weight	H_2 %	CO_2 %	H_2O %	CH_4 %	C_2H_6 %	C_3H_8 %	O_2 %	N_2 %
Propane-air-mixture	29.4	—	—	0.9	—	—	4.0	20.0	75.1
Propane-exhaust gas	28.2	—	11.6	16.6	—	—	—	—	71.8
Field gas[x]-air-mixture	26.0	8.5	1.2	1.1	4.7	1.1	—	17.5	65.9
Field gas[x]-exhaust gas	27.0	—	8.3	23.2	—	—	—	—	68.5

x: corresponds to III in Table 21.

Table 23. Physical properties of fuel-air mixtures and exhaust gases. (Same gases and conditions as in Table 22.) CGS-units.

~~Conversion factors: [illegible]~~

Temperature t		~~Specific weight, at 0°C, atmospheric pr. gram/Nm³~~	Specific heat, $\bar{c}_p$ Mean value between 0°C and t°C at constant pressure, cal/cm³ °C			Heat conductivity, λ at t°C 10^{-6} cal/cm, sec, °C.			Kinematic viscosity, ν at t°C and atmospheric pressure stokes (=cm²/sec)		
°C	°F		Propane-air-mixt.	Field gas air-mixt.	Exhaust gas	Propane-air mixt.	Field gas air-mixt.	Exhaust gas	Propane-air-mixt.	Field gas air-mixt.	Exhaust gas
0	32		0.325	0.313	0.327	53	81	48	0.09	0.18	0.11
100	212					68		65	.19	.34	.21
200	392		.336	.323	.335	82	127	79	.27	.52	.34
300	572					96		94	.39	.74	.49
400	752		.349	.330	.344	111	170	110	.50	.95	.63
500	932					125		126	.62	1.20	.77
600	1112		.361	.337	.353	139	210	142	.78	1.45	.97
700	1292					153		157	.92	1.76	1.16
800	1472		.377	.347	.362	166	249	172	1.08	2.04	1.3[illegible]
900	1652					179		186	1.24	2.33	1.55
1000	1832		.391	.353	.370	191	289	200	1.40	2.62	1.75
Conversion factors:			1 cal/cm³, °C = 0.0624 BTU/cu.ft, °F			1 cal/cm, sec, °C = 242 BTU/h, ft, °F			1 stoke = 0.001076 sq.ft/sec		

The exhaust gas from propane-air and from field gas-air have within ±1/2 % identical specific heat, heat conductivity and kinematic viscosity.
$\bar{c}_p$ valid for any constant pressure between 0–10 atm. λ independent of pressure,
ν inversely proportional to pressure.

Table 24. Gas densities at 32°F, atmospheric pressure.

Propane-air mixture:	1.31 grams/Nm³	= 81.4 lbs/10⁶ cuft
Propane-exhaust gas:	1.26 — " —	= 78.4 — " —
Field gas-air mixture:	1.16 — " —	= 72.2 — " —
Field gas-exhaust gas:	1.21 — " —	= 75.2 — " —

Table 25. Inflammability limits and ignition temperatures for gases in mixture with air at atmospheric pressures.

(For ignition temperatures, see also Table 28.)

Gas	Lower inflamm. limit % gas by volume = L_L	Upper inflamm. limit % gas by volume = L_U	Ignition temp. °F
Hydrogen	4.1	75	1075 – 1095
Methane	5.0	15	1200 – 1400
Ethane	3.0	14	970 – 1170
Ethylene	3.0	29	~1010
Propane	2.4	9.5	920
~~Heptane~~ Butane	1.9	8.5	765
Pentane	1.5	7.5	550
Carbon monoxide	12.5	75	~1190
Hydrogen sulphide	4.3	45	560
Field gas* (#III in Table 22)	4.6	28.3	

*) calculated with formula below.

Calculation of inflammability limits ~~mixtures~~ for gas mixtures.

(Some gases do not obey these equations.)

Lower limit for mixture $= \dfrac{p_1 + p_2 + p_3 + \cdots + p_a + p_b + p_c}{\dfrac{p_1}{L_{L1}} + \dfrac{p_2}{L_{L2}} + \dfrac{p_3}{L_{L3}} + \cdots}$

Upper limit for mixture $= \dfrac{p_1 + p_2 + p_3 + \cdots}{\dfrac{p_1}{L_{U1}} + \dfrac{p_2}{L_{U2}} + \dfrac{p_3}{L_{U3}} + \cdots}$

where $p_1, p_2, p_3 \cdots$ = % by volume of inflammable gases in mixture

$p_a, p_b, p_c \cdots$ = — " — of inert — " —

$L_{L1}, L_{L2} \cdots$ = lower limits for components

$L_{U1}, L_{U2} \cdots$ = upper — " —

Note: $p_a, p_b, p_c \cdots$ are included in lower limit formula only.

Table 26. Theoretical flame temperatures for methane-air and propane-air mixtures at atmospheric pressure. ~~(Centigrade)~~ (Degrees Kelvin)

Inlet temp. of mixture, °C	Methane-air					Propane-air				
	67%	83%	*100%	123%	150%	67%	83%	100%	123%	150%
25	1906	2122	2227	2025	1782	1975	2187	2267	2071	1822
50	1923	2138	2239	2042	1801	1992	2203	2279	2087	1840
75	1940	2155	2251	2058	1819	2009	2219	2290	2103	1858
100	1958	2172	2263	2074	1837	2027	2235	2301	2118	1877
150	1993	2205	2287	2106	1874	2062	2267	2324	2150	1913
200	2028	2238	2310	2138	1911	2098	2299	2346	2181	1950
250	2064	2270	2334	2170	1948	2134	2331	2369	2211	1986
300	210[illegible]	2303	2357	2201	1984	2171	2361	2391	2241	2023
350	2137	2336	2379	2232	2021	2207	2391	2413	2270	2059

100% corresponds to stoichiometric gas-air-mixture
123% — " — " 23% excess air.
(Source: U.S. Bureau of Mines)

Table 27. Ignition (= flame propagation) velocities for hydrogen, carbon monoxide, methane and propane in mixture with air at atmospheric pressure. (Gas mixtures not preheated.)

% air in mixture of stoichiometric	Ignition velocities, cm/sec.				
	Hydrogen	Carbon monoxide	Methane	Propane	Ethane
20	70	15	—	—	
30	195	33	—	—	
40	236	43	—	—	
50	265	48	—	—	
60	268	52	5	14	
70	253	50	16	22	
80	237	47	24	28	
90	218	44	26	30	
100 (= stoich.)	195	40	24	28	
110	168	37	20	24	
120	135	—	15	20	
Max. value, cm/sec	285	52	27	29	32
at % air	55	60	86	86	98

(Source: Corsiglia, Amer. Gas. Ass. M'ly, Oct. 1931, pp. 437–442.)

Ignition velocity = approximately proportional to T_2^a where T_2 ~~[illegible]~~ (abs. temp. of preheated gas-air-mixture) and a is slightly over 1.

Table 28. Ignition temperatures for air-gas mixtures

2% methane + 98% air	ignites at 1562°F	1.2% propane + 98.8% air ignites at	1090°F
4 " 96 "	" 1490	4.9 " 95.1 " "	977
8 " 92 "	" 1472	1.2% butane + 98.8% air "	1056
1.9% ethane + 98.1% air	" 1102	3.6 " 96.4 " "	959
8.1 " 91.9 "	" 1004	7.6 " 92.4 " "	912
6% ethylene + 94% air	" 1112	1% hydrogen sulphide + 99% air ign. at	703
10 " 90 "	" 1067	8 " 92 " "	582
25 " 75 "	" 1004		

(Increase in pressure lowers ignition temperatures. Substitution of air with oxygen also lowers ignition temperatures.)

Table 29. Effect of pressure on flammability limits for natural gas and methane ~~(in air)~~ (B.o.M. Information Circ. 7601.)

Gas	Pressure	Lower flamm. limit % gas by volume	Upper flammab. limit % gas by volume
Natural gas	70 mm Hg	4.4	11.2
	740 — " —	4.5	14.2
	500 psig	4.4	44.2
	3000 psig	3.15	60.0
Methane	42[illegible] mm Hg	no flammability	
	70 — " —	4.8	12.2
	760 — " —	5.0	15.0

Literature on ignition and explosion in gas-air mixtures.

1) Scott - Kennedy - Zabetakis: Gas explosions and their prevention. Bureau of Mines, Information Circular, Pittsburgh 1957.

2) Coward - Jones: Limits of inflammability of gases and vapors. B.o.M. Bulletin 279, Pittsburgh 1939. Revised in Bull. 503 (1952)

3) Jones - Kennedy - Spolan: Effect of high pressures on the flammability of natural gas - air - nitrogen mixtures. B.o.M. Report of Investigations 4557, Pittsburgh 1949

4) Henderson: Combustible gas mixture in pipe lines. 1941 Annual Proceedings of the Pacific Coast Gas Association.

5) Murphy: Rupture Diaphragms. Calculations, Characteristics and Uses. Chem. & Met. Eng. Nov/1944, pp. 108-112 and Dec. 1944, pp. 99-103.

6) Guest: Ignition of Natural Gas- Air Mixtures by Heated Surfaces. B. o. M. Techn. Paper 475, 1930, 59 pp.

~~Test~~ burner tests.

1-inch burner in 3-inch casing

46 ft burner tube, 50% sand and 59,000 BTU/h

(121-4B) gave a heated interval of ~70 ft and $\alpha_H = 93\%$, $\alpha_Z = 89\%$
(although a big sand loss!)

1-inch burner in 3½-inch casing

53 ft burner tube tested with 66,000 and 82,000 BTU/h

(122-B) Poor heat distribution and big sand loss.

1¼-inch burner in 3-inch casing

46 ft burner tube, 40% sand and 57,000 BTU/h gave

(121-2) a heated interval of ~70 ft and $\alpha_H = 105\%$; $\alpha_Z = 83\%$

36 ft burner tube, 50% sand and 57,000 BTU/h gave

(120-4) a heated interval of ~65 ft and $\alpha_H = 106\%$; $\alpha_Z = 86\%$

1¼-inch burner in 3½ inch casing

~~53~~ 53 ft burner tube tested with 59,000 and

(122-) 74,000 BTU/h. Poor heat distribution.

1½-inch burner in 4-inch casing 1120 BTU/ft

53 ft burner tube with 40% sand and 86,000

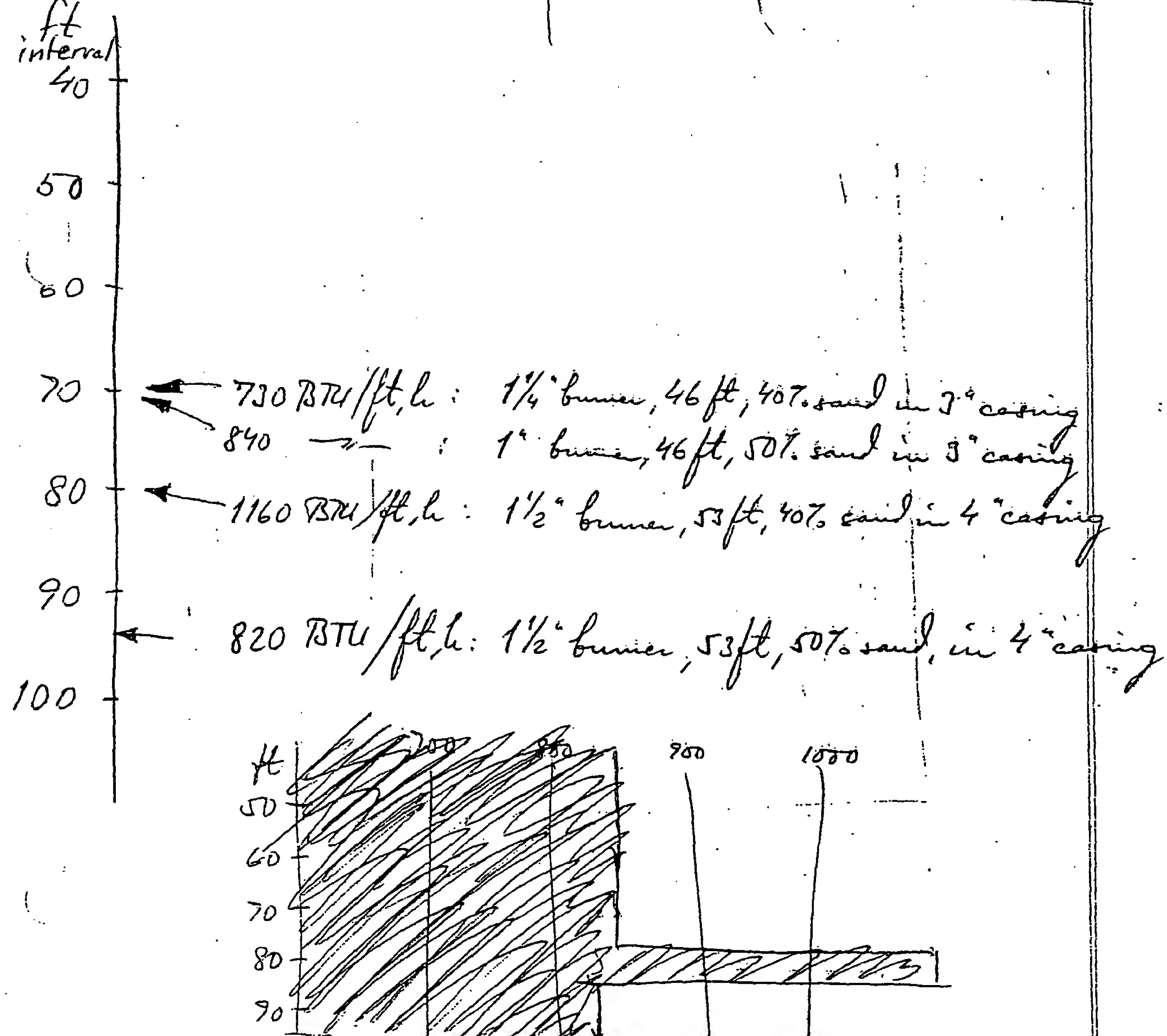
(126-2) BTU/h gave a heated interval of ~82 ft, $\alpha_H = 88\%$; $\alpha_Z = 83$

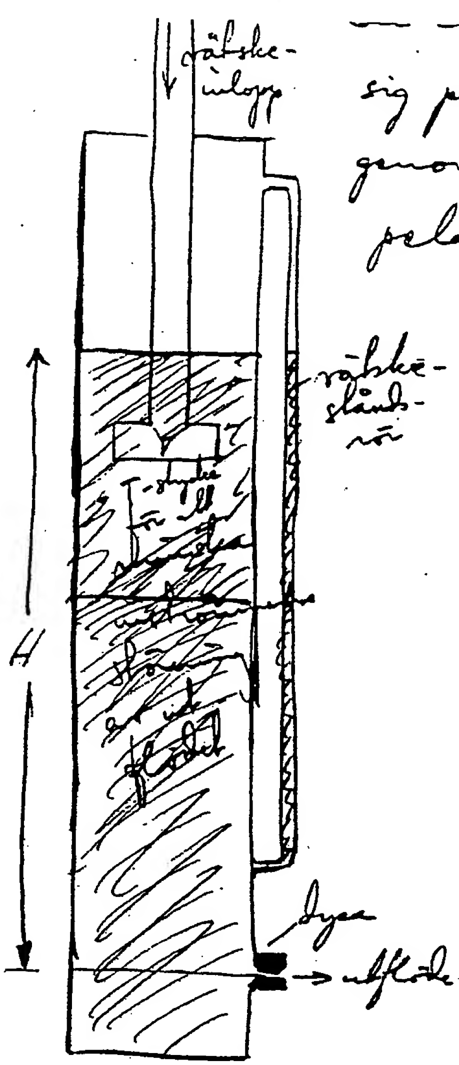
and 53 ft burner tube with 50% sand and 77,000

(126-3) BTU/h gave a heated interval of 94 ft and $\alpha_H = 86\%$, $\alpha_Z = 95\%$

820 BTU/ft

burner tube diam. / casing diam.	1"	1¼"	1½"
3"	70 ft 59,000 BTU	70 ft 57,000 BTU	X
3½"	poor	poor	X
4"	X	X	82 ft / 95,000 BTU 94 ft / 77,000 BTU





sig på tryckfallet vid en vätskas utströmning genom en dysa. Tryckfallet uppmätes som vätske-pelare, varvid den strömmande vätskan själv användes som manometervätska.

Utströmningen är då:

$$Q = K \cdot A \cdot \sqrt{2g \cdot H}$$

flöde — koeff. ca = 0.97 — dys-area — tyngd-kr. — vätske-pelarens höjd.

1.9.1959: Apparaten kan utformas praktiskt på olika sätt:

a) en, konstant dysa; gradering utefter vätskeståndsröret.

b) flera konstanta dysor, som öppnas eller stänges med kikkran; motsvarande antal skalor på vätskeståndsröret.

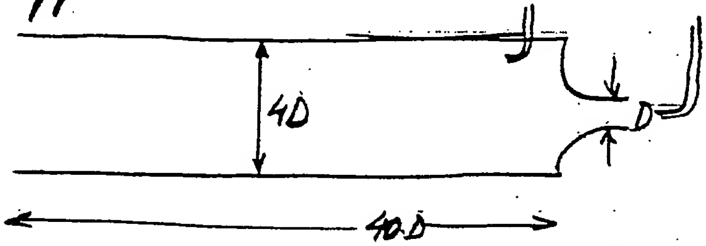
c) en variabel dysa av typ kamerabländare; ett märke på vätskeståndsröret. Dysan ändras tills ~~märk~~ vätskan in-ställt sig vid märket. Dysans inställning avläses.

Anordningen kan, t.ex. i utformning c) användas som doseringsanordning.

Sol

(Jmf. Compressed air data book.)

Mätningen sker med dysa med rundade kanter på inloppsidan och en diameter D i den parallella delen.



Dysan är insatt i änden av en 40D lång kammare med diametern 4D. Laminär utströmning förutsättes.

Tryckfallet över dysan och inloppsluftens tryck och temp. skall mätas.

Beteckningar:

Q_3 = luftflöde, cuft/minut vid trycket P_3 (= omgivande atmosfärens tryck) och temp. T_3 (= dito temp.)

D = dysdiameter, tum.

C = utströmningskoefficient, se tabell

T_1 = abs. temp, °F, före dysan

P_1 = abs. tryck, psia, —"—

P_2 = abs. tryck, psia, efter dysan

B = P_2, uttryckt i tum Hg (32°F)

I = tryckfallet $P_1 - P_2$, uttryckt i tum vatten.

$n = \frac{c_p}{c_v}$ för luft = 1.406 för torr luft
= 1.3947 för ~~fukt~~ luft med 36% fuktighet vid 20°C.

$r = \frac{P_1}{P_2}$;

$X = r^{0.283} - 1$

w = luftens täthet före dysan, dvs. vid P_1 och T_1.

Den teoretiska formeln (för adiabatisk strömning) är:

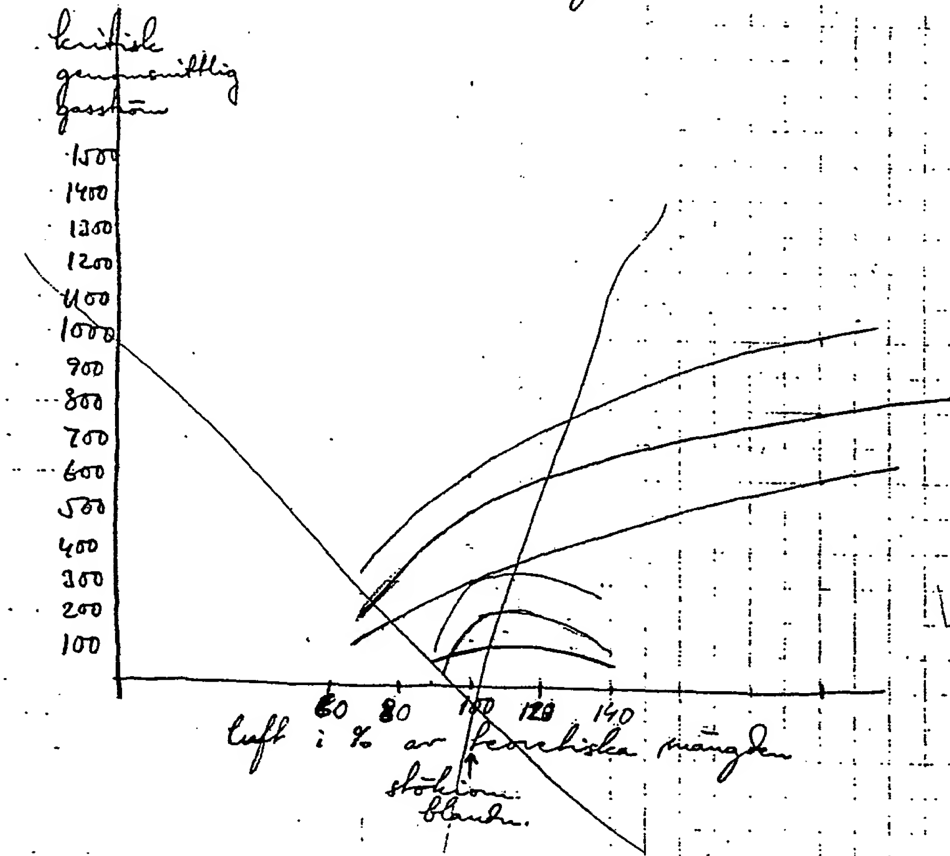
$$Q_3 = \frac{31,5 \cdot CD^2 \cdot P_2 \cdot T_3}{P_3 \cdot T_1 \sqrt{\frac{n-1}{n}}} \cdot \sqrt{\frac{P_1}{w}} \cdot \sqrt{\left(\frac{P_1}{P_2}\right)^{\frac{n-1}{n}} \left[\left(\frac{P_1}{P_2}\right)^{\frac{n-1}{n}} - 1\right]};$$

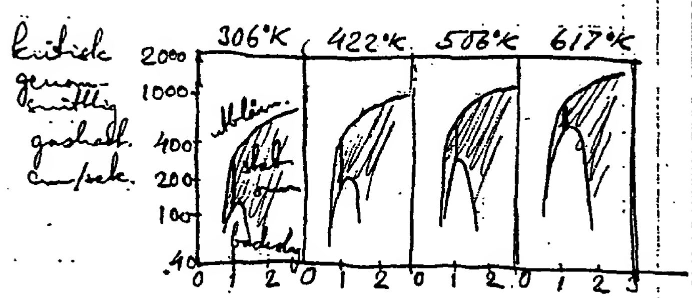
som kan skrivas: $Q = 59.22 \cdot \frac{CD^2 P_2 T_3}{\ldots} \cdot \sqrt{\frac{P_1}{w}} \sqrt{x(x+1)}$; ...

au:

$$Q_3 = \frac{17.16 \cdot D \cdot C \cdot T_3}{P_3} \cdot \sqrt{\frac{(P_1 - P_2) \cdot P_2}{T_1}};$$

Olika samband mellan gas-luftblandningens temperatur och lågans stabilitet har erhållits av olika forskare, men i allm. ~~[illegible] låga~~ står lågans vandringshastighet i proportion till T^a, där a är något större än 1 (kanske upp till 1.6).





Vidst. diagram visar hur stabilitetsområdet förflyttas vid stigande temp. på gas-luftblandningen.

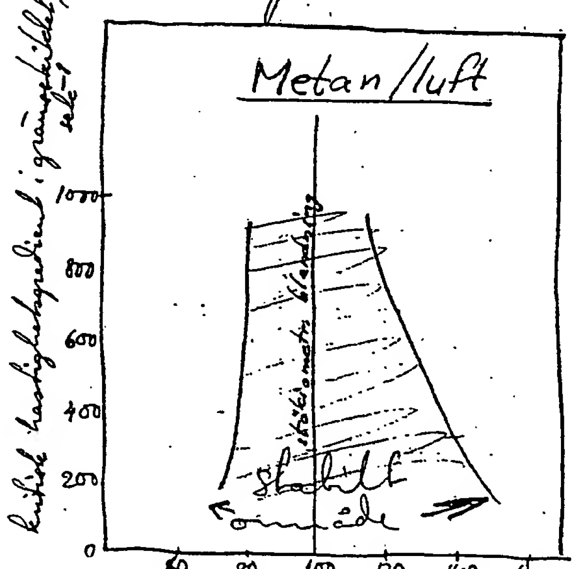
Sammandrag.

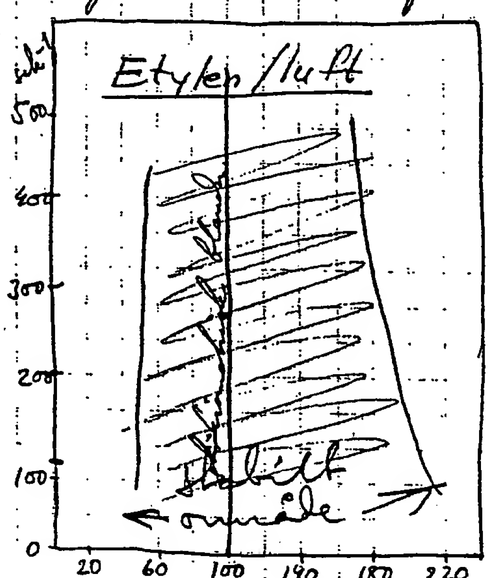
Vid försök i ett 150 cm långt rör, 5,0 cm diam. i vilket bränsle-luftblandningar fick brinna rätt-upp och nedåt, befanns att antändningsgränserna alltid äro vidare ~~vid upp flam uppåt strömmande gas~~ när lågan fortskrider uppåt än när den fortskrider nedåt, som tabellen visar: (röret slutet i antändningsändan).

	antändningsområde, % bränsle i bränsle-luftbland.	
	uppåtvandrande låga	nedåtvandrande låga
metan*	5.35 – 14.85	5.95 – 13.35
etan	3.15 – 14.8	3.32 – 10.0
n-pentane	1.43 – 8.0	1.49 – 4.56
etylen	3.13 – 33.3	3.42 – 15.3
propylen	2.21 – 9.6	2.29 – 7.2
acetylen	2.60 – 78	2.80 – 63.5
benzen	1.45 – 7.45	1.48 – 5.55
toluen	1.31 – 6.75	1.32 – 4.60

*) metanförsöken gjorda i 7.5 cm-rör ∅.

Om försök däremot göras i strömmande gas (ovanst. gjorda i vilande gasmassa), erhålles ingen skillnad i flam-stabilitet.





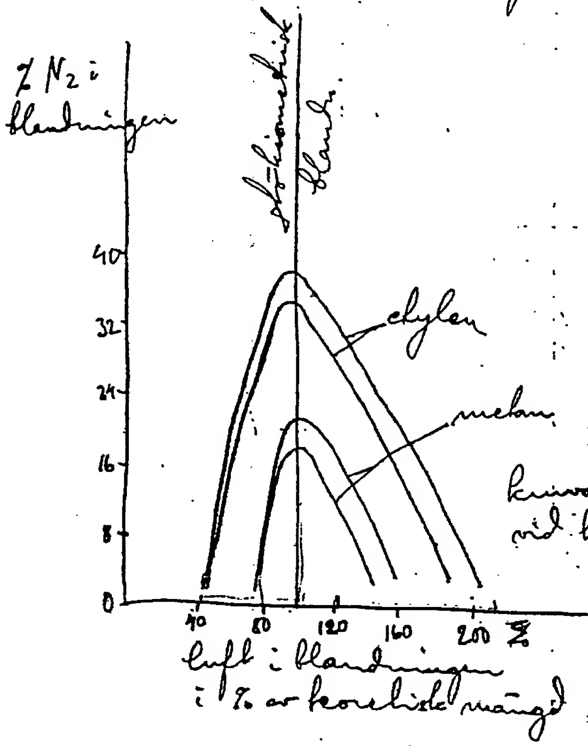
% luft i blandningen, i % av teoretisk luftmängd.

Diagrammen gälla för såväl uppåt- som nedåtgående låga

taget tunjära blandningslagar för såväl antändningsgränser som flam-stabilitetsgränser. (Vid den "rika" gränsen (= den med liten luftunderskott) kan möjligen viss inhibitionsverkan av en gas på en annan iakttagas.)

<u>Inverkan av utspädning med kväve</u>

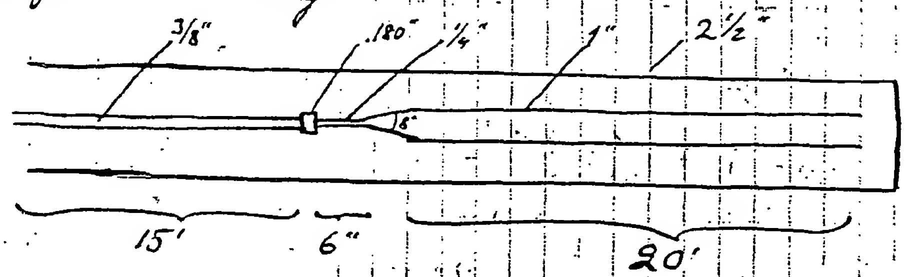
Till gas-luft-blandningar sattes kväve i varierande mängder.



Det befanns att vid den "magra" stabilitetsgränsen den enda effekten var en kylning. Vid den "rika" gränsen förekom möjligen någon quenching av kedjereaktionerna i flamman.

kurvorna erhållna vid två olika lufthastigheter

1. Tryckfallsberäkning



Pr timme tillföres brännaren 306,3 st/cuft = 8700 Nliter P-bränsle = 2420 Ncm³/sek.

Vid förbränningen därav bildas 8900 Nliter/h = 2480 Ncm³/sek P-rökgas.
Därav förutsättes att 25% = 620 Ncm³/sek cirkuleras via ejektorn.
Strömningarna skulle då bli: (tryck och temperaturer uppskattade).

sträcka	temp. °C	tryck ata	normal-volym Ncm³/sek	verklig volym cm³/sek	längden	id cm
medledn.-rör	50	1,5	2420 PB	1910	450	1,25
ejektor, .180"	50	1.5	2420 PB	1910	—	0,46
ejektorhals	100	1,2	2420PB+620PR	3460	15	0,92
kona, 8°	100	1,2	2420PB+620PR	3460	—	~~0,92~~ 2,4
brännarrör	800	1,1	3100 PR	11,100	600	2,66
ca 220 cm² nedre ringspalt (4 m = 2,9)	500	1,0	3100 PR	8.800	600	$\frac{d}{D} = \frac{3,32}{6,28}$
ca 285 cm² övre ringspalt (4 m = 4,5)	150	1,0	2480 PR	3.850	465	$\frac{d}{D} = \frac{1,92}{6,28}$

Följande Reynolds tal och ~~friktionsfaktor~~ hastigheter erhålles:

sträcka	ν stoke	R_e	hastighet cm/sek	f	anmärkn.
medledn. rör	0.093	~~481~~ 20.800	1550	0,0039	~~laminär~~ turbulent strömning
ejektor	0.093	~~(495)~~ 57.000	11.500	0,0035	—"—
ejektorhals	0,16	~~230~~ 30.000	5.200	0,0037	—"—
kona	0.16	—			
brännarrör	1.35	~~7200~~ 3950	2.000	0,0054	turbulent strömning
plötslig utvidgn.	1.35	—			
nedre ringsp.	0.77	~~900~~ 1570	400	0,008	laminär strömning
övre ...		2300			

~~nedledningsröret: $\Delta p = \frac{\quad}{g \cdot D^2} = 0{,}00505 \cdot L \; g/cm^2$~~

~~För ett 450 cm långt rör blir tryckfallet alltså 2,3 g/cm^2.~~

$$\Delta p = 8 \cdot F \cdot \frac{L}{D} \cdot \frac{\rho \cdot v^2}{2g} = \quad \cancel{= 0.033 \text{ psi}}.$$

$$= 8 \cdot 0{,}0039 \cdot \frac{450}{1{,}25} \cdot \frac{0.00173 \cdot 1550^2}{2 \cdot 981} = 0{,}0525 \cdot 450 = \underline{\underline{23{,}6\ g/cm^2}}$$

$$= \underline{\underline{0{,}35 \text{ psi}}}.$$

ejektorn:

behandlas som en strypning med en area $= \left(\frac{0{,}46}{1{,}25}\right)^2 = 0{,}135$ av nedledningsrörets area; enl. TDF är S då = ca 2,3 velocity heads, eller $L_e = D \cdot \frac{2{,}3}{8 \cdot F} = 1{,}25 \cdot \frac{2{,}3}{8 \cdot 0{,}0039} = 92{,}2$ cm, motsvarande ett tryckfall genom ejektorn av $\frac{92}{450} \cdot 23{,}6 = \underline{\underline{4{,}82\ g/cm^2}}$.

ejektorhalsen:

$$\Delta p = 8 \cdot 0{,}0037 \cdot \frac{15}{0{,}92} \cdot 0{,}00114 \cdot \frac{5200^2}{2 \cdot 981} = \underline{\underline{7{,}6\ g/cm^2}}.$$

konan: $S = 0{,}06$ velocity heads, d.v.s. $L_e = D \cdot \frac{0{,}06}{8 \cdot F} = 0{,}92 \cdot \frac{0{,}06}{8 \cdot 0{,}0037} =$

$= 1{,}9$ cm, motsvarande ett tryckfall av $\frac{1{,}9}{15} \cdot 7{,}6 = \underline{\underline{0{,}95\ g/cm^2}}$

brännarröret: $\Delta p = 8 \cdot 0{,}0054 \cdot \frac{600}{2{,}66} \cdot \frac{0{,}000352 \cdot 3950^2}{1962} = \underline{\underline{27{,}2\ g/cm^2}}$

plötslig övergång: här blir $S = \left(1 - \frac{25{,}56}{22{,}0}\right)^2 = 0{,}56$, d.v.s.

$L_e = D \cdot \frac{0{,}56}{8 \cdot F} = 2{,}66 \cdot \frac{0{,}56}{8 \cdot 0{,}0054} = 34{,}5$ cm. d.v.s.

$\Delta p = \frac{34{,}5}{600} \cdot 27{,}2 = \underline{\underline{1{,}6\ g/cm^2}}.$

nedre ringspalten: $\Delta p = 8 \cdot 0{,}008 \cdot \frac{600}{2{,}91} \cdot \frac{0{,}0044 \cdot 16 \cdot 10^4}{1962} = 0{,}475\ g/cm^2$

övre ringspalten:

$$\Delta p = 8 \cdot 0{,}005 \cdot \frac{465}{4{,}55} \cdot \frac{0{,}00081 \cdot 136^2}{1962} = \underline{\underline{0{,}031\ g/cm^2}}$$

alltså totalt $\Delta p = 4{,}8 + 23{,}6 + 7{,}6 + 1{,}0 + 27{,}2 + 0{,}5 = 65\ g/cm^2 = 0{,}95$ psi.

2. Värmebalans.

Per timme tillföres brännaren 8,70 Nm^3 propan-luftblandning och bortgår i form av "propanrökgas" 8,90 Nm^3. I brännarens olika delar förekomma dessa gaser vid olika temperaturer. Deras värmeinnehåll har nedan beräknats med ledning av tabellen över medelspecifika värmet.

temp. °C		bränslegasens värmeinnehåll kcal/h	BTU/h		rökgasens värmeinnehåll kcal/h	BTU/h
0	0,325	—	—	325	—	—
100	330	287	1140	330	294	1165
200	336	585	2320	335	596	2360
300	342	895	3540	339	905	3580
400	349	1220	4840	344	1220	4840
500	355	1550	6150	348	1550	6150
600	361	1880	7460	353	1890	7500
700	369	2240	8.900	358	2230	8850
800	377	2630	10.400	362	2580	10.250
900	384	3060	12.150	366	2940	11.650
1000	391	3400	13.450	371	3300	13.100
1100	398	3810	15.100	376	3690	14.600

Gissade temperaturer.

Noggrannare temperaturuppskattningar komma senare att göras. Här antages preliminärt följande temperaturförhållanden i brännaren:

a) bränsletillförselrörets övre ände: 20°C ∴ gasens fysiska värme ~60 kcal/h

b) ~~[illegible]~~ inträdet i ejektorn: 100°C ∴ bränslegasens — " — ~290

c) återförd rökgas (25%): 200°C ∴ återförda gasens — " — ~150

d) bränsle och återförd rökgas blandas adiabatiskt och införes i konan, där förbränning sker och 30.000 BTU/h = 7550 kcal/h frigöres. Till förfogande står då 7550 + 290 + 150 = 7990 kcal/h.

e) rökgasen (+25%) lämnar brännarens hela zon (konan + 2 à 3 fot rör) vid en temperatur - - 700, 800, 900, 1000, 1100°C. Dess h[illegible]

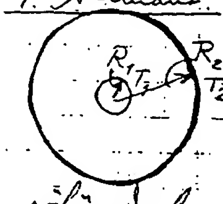
~~[illegible]~~ sugningsöppningarna för ejektorn. Dess fysiska värme-innehåll är då ca. ~~600~~ 750 kcal/h, varav ~~[illegible]~~ 150 återgår till brännaren, medan resten, ~~[illegible]~~ 600 kcal/h, går ut med rökgasen. På vägen upp avges en del värme till bränslegasen och till ovan-bädden-lagren. Dess temperatur sjunker därvid till ca. 100°C omedelbart före utträdet i det fria, då dess värmeinnehåll alltså är ca. 300 kcal/h.

Det framgår bl.a. av detta överslag att det värme, som avges genom strålning från brännarens hetaste del är av storleksordningen 7990 − (3000 à 4000) kcal/h eller ungefär hälften av den netto-tillförda effekten.

Värmeövergång:

Värmet frigöres ur bränslet i brännarörets övre del, där förbränningen sker. Här antages t. v. att förbränningen sker i konan och brännarörets översta en meter långa del. Den per timme tillförda värmemängden är 7990 kcal (jfr. avsnitt 2). Vid jämvikt skall denna effekt bortföras, vilket sker <u>dels</u> genom strålning ~~dels genom konvektion från rörväggen till rökgasen på utsidan,~~ (som antas ske enbart radiellt utåt), <u>dels</u> genom axiell ledning i rörväggen (så ringa att den t. v. försummas) och <u>dels</u> genom axiell utströmning av het rökgas.

1. Allmänt   Värmeöverföring genom strålning.



Värmestrålningen från det inre, heta röret till det yttre, svala röret är:

rörlängd = l

$$H = c\,(2\pi R_1 l T_1^4 - 2\pi R_2 l T_2^4) \qquad (1)$$

där konstanten c är ca $1\cdot10^{-12}$ cal·sek^{-1}·cm^{-2}·°C^{-4}

d.v.s.

$$\frac{H}{l} = 6{,}28\cdot10^{-12}\,(R_1T_1^4 - R_2T_2^4)\ \text{cal sek}^{-1}\ \text{cm}^{-1} \qquad (2)$$

2. Rördimensionernas inverkan

Ekvation (2) visar att om en viss värmemängd skall tillföras ett beklädnadsrör med given dimension och temperatur, så blir bränsleröets temperatur högre ju mindre det är.

Exempel. Antag att beklädnadsröret är 4" d.v.s. $R_2 = 5{,}0$ cm och att det är gjort av material som inte tål högre temperaturer än 800°K (för att värmeöverföringen skall bli så effektiv som möjligt bör högsta tillåtna temperatur på beklädnadsröret utnyttjas). Den överförda effekten antages vara 3 cal·cm^{-1}·sek^{-1} (motsvarande 10.800 kcal/m timme).

Vi få alltså

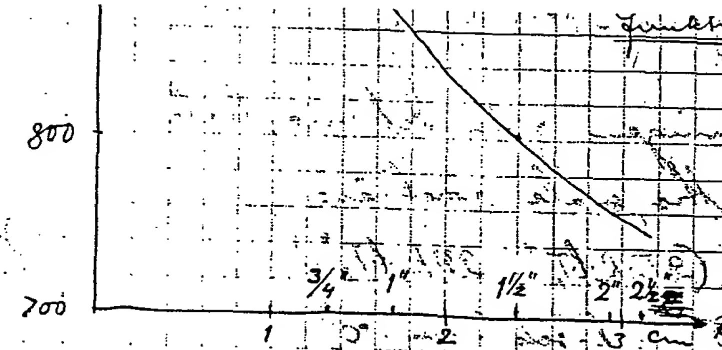
$$3 = 6{,}28\cdot10^{-12}\,(R_1T_1^4 - 5\cdot800^4) \qquad (3)$$

~~$0{,}48\cdot10^{12} + 0{,}41\cdot10^{12} = R_1T_1^4$~~

$$R_1T_1^4 = 0{,}99\cdot10^{12} \qquad (4)$$

Vid olika dimensioner av innerröret erhållas följande temperaturer:

nominell rörstorlek	R_1 cm	temperatur T_1 °K
1/2 tum	1,07	981
3/4 "	1,33	930
1 "	1,67	877
1 1/2	2,40	801



3. ~~…~~ Inverkan av en strålningsskärm

Om den överförda värmemängden är för stor, kan strålningen minskas genom insättandet av en (eller flera) skärm mellan inner- och ytterrör (~~…~~ koncentriskt rör)

En skärmradie R'

från innerrör till skärm: $$\frac{H}{l} = c\,2\pi\left(R_1 T_1^4 - R'(T')^4\right) \quad (5)$$

och från skärm till ytterrör: $$\frac{H}{l} = c\,2\pi\left(R'(T')^4 - R_2 T_2^4\right) \quad (6)$$

ty intet värme ackumuleras i skärmen.

Alltså: $$R_1 T_1^4 - R'(T')^4 = R'(T')^4 - R_2 T_2^4 \quad (7)$$

eller

$$R'(T')^4 = \frac{R_1 T_1^4 + R_2 T_2^4}{2} \quad (8)$$

och

$$\frac{H}{l} = c\cdot 2\pi\left(\frac{R_1 T_1^4 - R_2 T_2^4}{2}\right) \quad (9)$$

Dvs. den strålade värmemängden blir blott hälften så stor som den skulle ha varit, om skärmen ej funnits.

P.s.s. kan visas att två skärmar minskar strålningen till en tredjedel, tre skärmar till en fjärdedel osv.

Obs! Dessa samband förutsätter att temperaturerna T_1 och T_2 är desamma med och utan skärm. Emellertid betyder insättandet av skärmen och därmed minskningen av värmeöverföringen att innerrörets temperatur stiger och ytterrörets sjunker. Om värmefältet kan avledas åt innerhålle

Net radiation heat transfer in annuli

$$q_{rad.} = \frac{A_1}{\frac{1}{\alpha_1} + \frac{A_1}{A_2}\left(\frac{1}{\alpha_2} - 1\right)} \cdot \sigma \cdot (T_1^4 - T_2^4)$$

A_1 och A_2 = resp. rör ytor

α_1 och α_2 = emissivitetskonstanter

σ = Stefan-Boltzmann ~~lag~~ konstant

G. P.: $P = 0,173 \cdot \varepsilon \cdot A \cdot F\left[\left(\frac{T_1}{100}\right)^4 - \left(\frac{T_2}{100}\right)^4\right]$

ε = helizyn = 0.70

$T_1 = \sim 1050\,°C$; $T_2 = \sim 700\,°C$

Värmeöverföring vid både in- och utsidan

Innerröret: r_1, t_1, q_1 (= $P_1 \cdot 100$ % av q_1+q_2) $q_1+q_2=q$

Ytterröret: r_2, t_2, q_2 (= $P_2 \cdot 100$ % — —)

Δt = temperaturfallet i vätskan över rörets längd L.

$$\left.\begin{aligned} q_1 &= P_1 q = -k_1 \cdot 2\pi r_1 \left(\frac{\partial t}{\partial r}\right)_1 \\ q_2 &= P_2 q = k_2 \cdot 2\pi r_2 \left(\frac{\partial t}{\partial r}\right)_2 \end{aligned}\right\}$$

Insättes i den allmänna diff-ekvationen för annulär och löses, erhålles:

$$\frac{t_2-t_1}{\Delta t} = \frac{V \cdot C_p}{2\pi k \cdot L} \cdot X_1$$

V = flödet i cm³/sek

$C_p = \rho \cdot c_p$ = cal/°C, cm³.

där X är en formfaktor, som dels beror av $\frac{r_1}{r_2}$, dels beror av P_1.

$\frac{q_1}{q_1+q_2}$ \ $\frac{r_1}{r_2}$ =	0	0,01	0,025	0,05	0,1	0,2	0,4	0,6	0,8	1,0
[illegible]	$X=\infty$	1,602	1,395	1,229	1,041	0,804	0,524	0,325	0,157	0
0	0.75	0,681	0,658	0,629	0,581	0,490	0,343	0,221	0,106	0
0,2	$-\infty$	−0,240	−0,080	0,029	0,121	0,170	0,162	0,117	0,0612	0
0.4	$-\infty$	−1,161	−0,818	−0,571	−0,339	−0,150	−0,019	0,013	0,0164	0
0,6	$-\infty$	−2,082	−1,556	−1,171	−0,799	−0,470	−0,200	−0,091	−0,0289	0
0,8	$-\infty$	−3,003	−2,294	−1,771	−1,259	−0,790	−0,381	−0,195	−0,0732	0
1,0	$-\infty$	−3,924	−3,032	−2,371	−1,719	−1,110	−0,562	−0,299	−0,118	0
1,2	$-\infty$	−4,845	−3,770	−2,971	−2,179	−1,430	−0,743	−0,403	−0,163	0

Strömning i annuli.

Diametrar d_1 och d_2 ; $\frac{d_1}{d_2} = y$.

Laminär str.		Turbulent str.	
släta rör	handelsrör	släta rör	handelsrör

$$\left.\begin{array}{ll} \text{för } 0<y<0{,}85 & j = \dfrac{8\mu\cdot d_2}{\mu\cdot\rho\cdot(1-y)}\cdot\dfrac{(1-y)^2}{1+y^2+\dfrac{1-y^2}{\ln y}} \\ \text{för } 0{,}15<y<0{,}85 & j = \dfrac{8\cdot\mu\cdot d_2}{\mu\cdot\rho\cdot(1-y)}\cdot 1{,}50 \\ \text{för } 0{,}85<y<1{,}00 & j = \dfrac{12}{Re} \end{array}\right\} \quad j = 0{,}004$$

Strömning i rör

För $2000 < Re < 50.000$:

$j = \dfrac{K}{Re^{0,25}}$; $K = 0{,}04$ för släta rör

$K = 0{,}027$ för handelsrör.

Konvektionsöverföring av värme i annuli.

(enl. Monrad-Pelton, Trans. AICE, 1942, pp 593-611)

(Experimentell undersökning)

Om en vätska strömmar i en annulus är hastighetsgradienten och därmed värmeöverföringen mycket större vid innerrörets yta än vid ytterrörets.

Dittus-Boelter's ekvation för rör:

$$\frac{h\cdot D}{k} = 0{,}0225\left(\frac{D\cdot u\cdot \rho}{\mu}\right)^{0,8}\left(\frac{c\cdot \mu}{k}\right)^{n}$$

där $n = 0{,}4$ för uppvärmning och $= 0{,}3$ för avkylning.

Ekvationen kan användas vid den yttre annulus-ytan om D ersättes med den ekvivalenta diametern $D_1 - D_2$.

Vid den inre annulus-ytan kan samma ekvation användas i modifierad form: $\left(\frac{D_2}{D_1} = Y\right)$

$$\frac{h\cdot(D_1-D_2)}{k} = 0{,}0225\cdot\frac{2\cdot\ln\frac{1}{Y} - \frac{1}{Y^2} + 1}{\frac{1}{Y^2} - Y - \frac{2}{Y}\ln\frac{1}{Y}}\cdot\left[\frac{(D_1-D_2)\cdot u\cdot \rho}{\mu}\right]^{0,8}\cdot\left(\frac{c\cdot \mu}{k}\right)^{n}$$

Tryckfall i annuli

Fannings ekvation för rör:

$$\Delta p = \frac{2\cdot f\cdot L\cdot \rho\cdot u^2}{g\cdot D} \quad \text{där} \quad f = 2\cdot 0{,}023\cdot\left(\frac{D\cdot u\cdot \rho}{\mu}\right)^{-0,2}$$

kan även användas för annuli, om faktorn 0,023 multipliceras med korrektionsfaktorn x:

$$x = \frac{\frac{1}{Y} + \frac{2\ln\frac{1}{Y} - \frac{1}{Y^2} + 1}{\frac{1}{Y} - Y - \frac{2}{Y}\ln\frac{1}{Y}}}{\frac{1}{Y} + 1},$$

Två koncentriska rör.

Det inre rörets ytterradie $= r_1$ och dess temp. $= t_1$.
Det yttre rörets innerradie $= r_2$ —"— $= t_2$
Kanalen (rörens) längd $= L$

1. Värme tillföres utifrån. Inget värme bortföres genom inre röret:

Det av fluidet upptagna värmet är då:

$$q = h_{21} \cdot 2\pi \cdot r_2 \cdot L \cdot (t_2 - t_1)$$

där $q =$ cal/sec

$h_{21} = k \cdot F_{21}$, där $k =$ fluidets värmeledningsförmåga
och $F_{21} =$ en geometrisk formfaktor, som är en funktion av $\frac{r_1}{r_2}$; (se tabell 1)

2. Värme tillföres inifrån. Inget värme bortföres från yttre röret.

Det av fluidet upptagna värmet är då:

$$q = h_{12} \cdot 2\pi \cdot r_1 L \cdot (t_1 - t_2)$$

där $q =$ cal/sec.

$h_{12} = k \cdot F_{12}$, där $F_{12} =$ en geometrisk formfaktor, som är en funktion av $\frac{r_1}{r_2}$; (se tabell 1)

Tabell 1

$\frac{r_1}{r_2}$	$\frac{h_{21} \cdot r_2}{k}$	$\frac{h_{12} \cdot r_1}{k}$
0.1	1,72	0.58
.2	2.01	0.90
.3	2,37	1.28
.4	2.85	1.77
.5	3.52	2.44
.6	4.52	3.45
.7	6.19	6,12
.8	9.52	8.47
.9	19.57	18.46

Vid driftfallet 30.000 BTU/20 ft, hour erhålles:

$$3{,}48 = 7{,}11 \cdot 10^{-12} (T_I^4 - T_Y^4) \quad \text{eller} \quad T_I^4 = 0{,}49 \cdot 10^{12} + T_Y^4.$$

$T = 540°C$	∴ vid $\tau = 14$ tim.	$T_Y^4 = 860 \cdot 10^8 = 0{,}086 \cdot 10^{12}$ och	$T_I^4 = 0{,}\cancel{402}\,576 \cdot 10^{12}$ då	$T_I = \cancel{880}\ 870$
595	28	$1250 \cdot 10^8 = 0{,}1250 \cdot 10^{12}$	$\cancel{0{,}\ldots}\ 0{,}615 \cdot 10^{12}$	$T_I = 887°$
720	140			935°
780	280			967°
900	1400			1038
960	2800	$0{,}846 \cdot 10^{12}$		1080

~~Strålningsskyddsinsats:~~
~~Om ett strålningsskydd insättes i ovannämnda brännare, inträffar ett nytt driftfall.~~

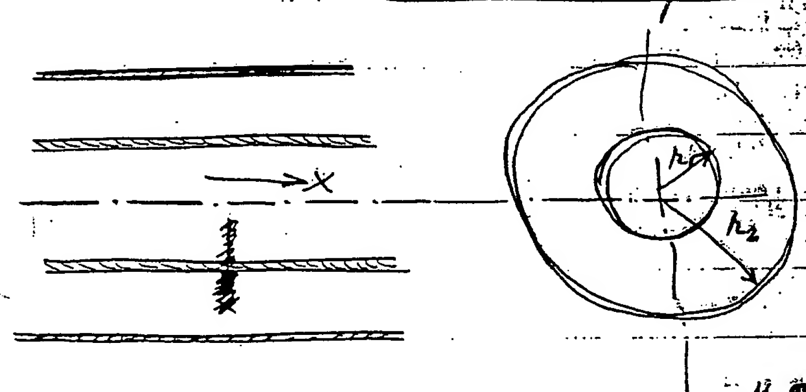
Dessa temperaturer T_Y och T_I äro riktiga endast i första approximationen. En del av det vid förbränningen frigjorda värmet "förloras" nämligen genom den från brännarröret utströmmande rökgasen (här bortses från att en del av detta värme kommer bruget tillgodo). Den ~~...~~ förlorade värmemängden kan antas = rökgasens fysiska värme vid brännarrörets temperatur. Den blir sålunda enl. ovan:

vid $\tau = 14$ tim.	$T_I = 870$; $\Delta Q = 870 \cdot 0{,}365 \cdot 8{,}9$ kcal/h $= 2820$ kcal/h	$= 11.200$ BTU/h
28		11.400 "
140		12.100 "
280		12.600 "
1400		13.600 "
2800		14.200 "

Anteckningar betr. gasflammor.

1. Om en gas får utströmma uppåt i luft ur ett vertikalt rör, bestämmes lågans geometri av diffusionsförloppet. Lågan är i stort sett en buktig yta av liten tjocklek. Dess höjd beror dels på gas-slaget (en CO-flamma är 2,5 ggr så hög som en H_2-flamma) dels på utströmningshastigheten, men däremot icke på bränslets grad av förvärmning.

Heat transfer to a fluid in laminar flow through an annular space.



Strömningen i ringspalten antages vara laminär och "steady state".
μ = vätskans dynam. viskos.
k = -"- värmeledn. förm.
μ och k antas temperaturoberoende.

Strömningen är = $V = \frac{\pi}{8\mu}\cdot M\cdot\frac{dp}{dx}$;

där $M = (r_2^2 - r_1^2)(r_2^2 + r_1^2 - B)$, där $B = \frac{r_2^2 - r_1^2}{\ln\frac{r_2}{r_1}}$;

Värmeströmningens allmänna differentialekvation är:

$$\frac{\partial}{\partial r}\left(r\frac{\partial t}{\partial r}\right) = N\cdot r\cdot\left(r_2^2 - r^2 + B\cdot\ln\frac{r}{r_1}\right)\frac{\partial t}{\partial x} - r\,\frac{\partial^2 t}{\partial x^2};$$

där $N = \frac{2V\cdot C_p}{\pi\cdot M\cdot k}$ = konstant

Fall 1: Jämn värmetillförsel från utsidan och fullst. isolering på insidan (t.ex. om innerröret utgöres av en solid stav, ett termoelement el. dyl.). Ekvationen blir i detta fall:

Det värme, som överföres genom yttre rörväggen på längden Δx är:

$$\cancel{q_2 = k \cdot 2\pi r_2 \left(\frac{\partial\theta}{\partial r}\right)_{r=r_2} \cdot \Delta x;}$$

$$\cancel{\text{där } \frac{\partial\theta}{\partial r} = N \cdot C \cdot \left[\frac{r_1^2 r}{2} - \frac{r^3}{4} + \frac{B r}{2} \cdot \ln\frac{r}{r_1} - \frac{r_1^4}{4r} + \frac{B r_1^2}{4r}\right]}$$

$$q_2 = h_{21} \, \Delta x \cdot 2\pi r_2 \cdot (t_2 - t_1)$$

där h_{21} = den sammansatta yt-överföringskoefficienten, som beräknas så:

$$h_{21} = -k \cdot \frac{\frac{r_1^2 r_2}{2} - \frac{r_2^3}{4} + \frac{B r_2}{2} \cdot \ln\frac{r_2}{r_1} - \frac{r_1^4}{4 r_2} + \frac{B r_1^2}{4 r_2}}{\cancel{\frac{4 r_1^4}{16}} - \frac{r_1^4}{16} + \cancel{\frac{B r_1^2}{4} \ln\frac{r_1}{r_1}} - \frac{B r_1^2}{4} + \left(\frac{r_1^4}{4} - \frac{B r_1^2}{4}\right)\ln\frac{r_2}{r_1} - \left(\frac{r_1^2 r_2^2}{4} + \frac{r_2^4}{16} - \frac{B r_2^2}{4} \ln\frac{r_2}{r_1} + \frac{B r_2^2}{4}\right)}$$

Men $$h_{21} = -k \cdot \frac{2 r_1^2 r_2 - r_2^3 + 2B r_2 \ln\frac{r_2}{r_1} - \frac{r_1^4}{r_2} + B \cdot \frac{r_1^2}{r_2}}{\frac{3 r_1^4}{4} - B \cdot r_1^2 + B \cdot r_2^2 - r_1^2 r_2^2 + r_2^4 - \left(\ln\frac{r_2}{r_1}\right) B \left(r_1^2 + r_2^2\right)}$$

men $B = \frac{r_2^2 - r_1^2}{\ln\frac{r_2}{r_1}}$; $\quad B \cdot \ln\frac{r_2}{r_1} = (r_2^2 - r_1^2)$

$$\therefore h_{21} = -k \cdot \frac{2 r_1^2 r_2 - r_2^3 + 2 r_2^3 - 2 r_2 r_1^2 - \frac{r_1^4}{r_2} + \frac{r_1^2}{r_2} \cdot \frac{r_2^2 - r_1^2}{\ln\frac{r_2}{r_1}}}{\frac{3 r_1^4}{4} - \frac{(r_2^2 - r_1^2)^2}{\ln\frac{r_2}{r_1}} - r_1^2 r_2^2 - r_2^4 - \frac{r_2^4 - r_1^4}{\ln\frac{r_2}{r_1}}} = \frac{T}{N}$$

$$-\frac{1}{\ln} \cdot \left[r_2^4 + \cancel{r_1^4} - 2 r_1^2 r_2^2 + r_2^4 - \cancel{r_1^4}\right] = -\frac{1}{\ln} \cdot \left[2 r_2^4 - 2 r_1^2 r_2^2\right] = + \frac{2 r_2^2}{\ln}(r_1^2 - r_2^2)$$

$$\therefore h_{21} = -k \cdot \frac{T}{\frac{3 r_1^4}{4} - r_1^2 r_2^2 - r_2^4 + \frac{2 r_2^2}{\ln\frac{r_2}{r_1}}\left(r_1^2 - r_2^2\right)}$$

Vid isotermisk gasströmning gäller:

$$\Delta p = 8 \cdot \left(\frac{R}{\rho \cdot v^2}\right) \cdot \frac{L}{D} \cdot \frac{\rho \cdot v^2}{2g} \qquad (1)$$

där Δp = tryckfallet i g/cm² (vid stora tryckfall ersättes Δp med $\frac{p_1^2 - p_2^2}{2p_m}$)

L = ledningslängden, cm

D = ledningsdiametern, cm

$\frac{R}{\rho v^2}$ = friktionsfaktorn, som erhålles ur Reynolds tal, R_e

ρ = gasens täthet, g/cm³

v = gasens hastighet, cm/sek

g = 980,7 cm/sek²

Vid __laminär__ strömning ($R_e < 2100$) är friktionsfaktorn

$$\frac{R}{\rho \cdot v^2} = \frac{8}{R_e} \qquad (2)$$

Vid __turbulent__ strömning ($R_e > 2100$) erhålles friktionsfaktorn ur diagram.

__Reynolds tal__ = $R_e = v \cdot D \cdot \nu$

där ν = gasens kinematiska viskositet i stoke.

För __icke-cirkulära tvärsnitt__ ersättes D i formlerna med $4 \cdot m$

där $m = \frac{\text{tvärsnittsarean}}{\text{perimetern}}$;

För __laminär__ strömning i spalten mellan två koncentriska rör

gäller: $\Delta p = 8 \cdot \frac{L}{D} \cdot \frac{4 \cdot \eta \cdot v}{g \cdot D \cdot \left[1 + \alpha^2 + \frac{1-\alpha^2}{\ln \alpha}\right]}$;

där $\alpha = \frac{d}{D}$; och $\eta = \rho \cdot \nu$;

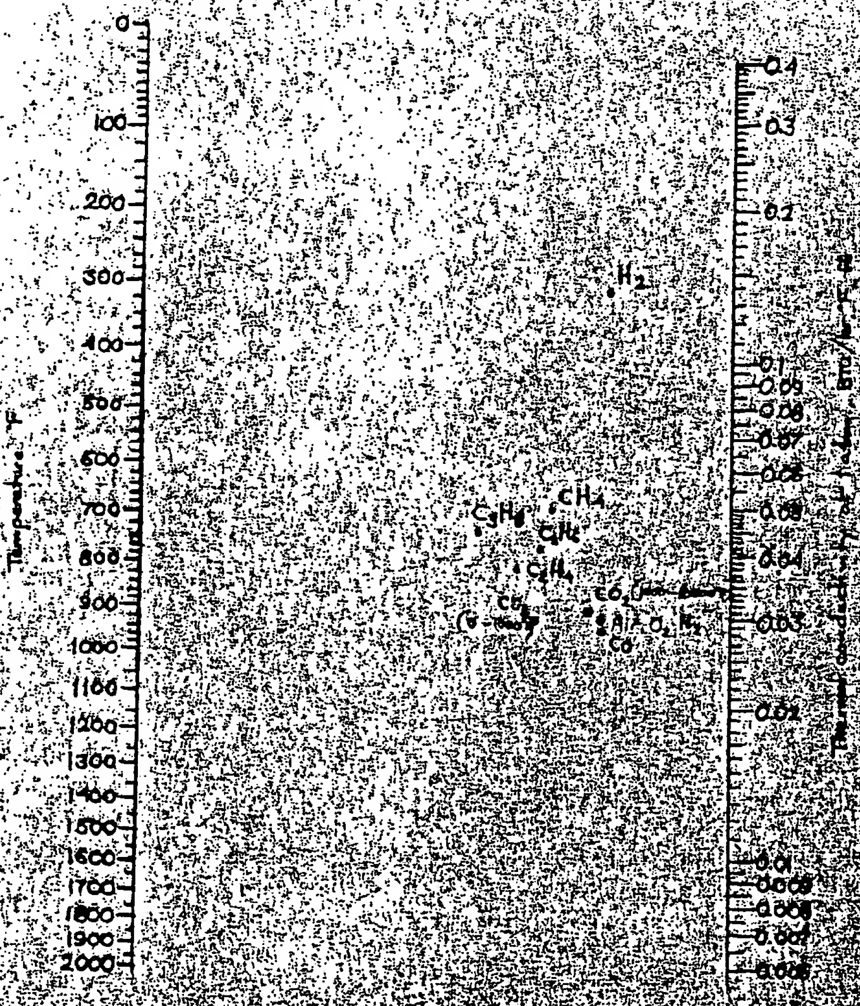
Fluid flow through packed and fluidized systems.
by M. Leva, M. Weintraub, M. Grummer, - - - (1951)

Pressure drop through fluidized beds.

$$\Delta P = \frac{V_t}{A_t}(1-\delta)(\rho_s-\rho)$$

where ΔP = pressure drop
V_t =
A_t =
δ = voids in beds
ρ_s = density of solid
ρ = " fluid

Thermal Conductivity Chart for Gases



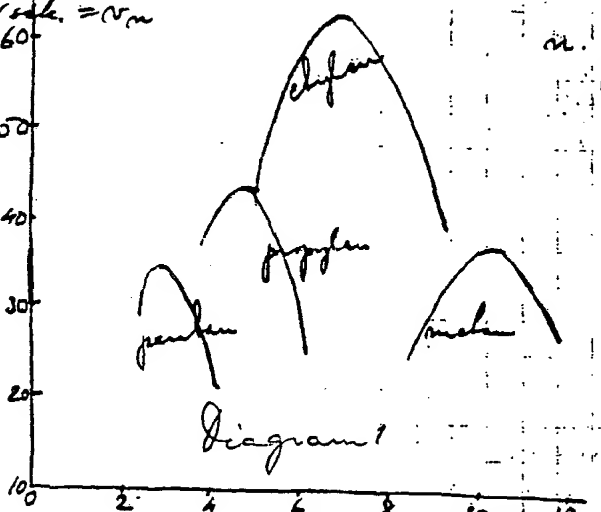
Man skiljer mellan: förbränning, detonation och explosion.
För LINS-brännaren är endast förbränning aktuell.

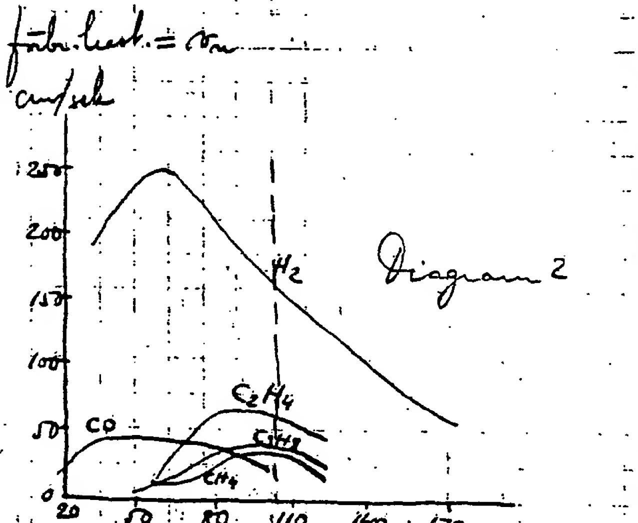
1. Tändtemperatur:

Luft, innehållande

2 %	metan tänds vid	850 °C
4	"	810
8	"	800
1,97 %	etan	594
8,15	"	540
1,25 %	propan	588
4,90	"	525
1,25 %	butan	569
3,65	"	575
7,65	"	489
6 %	etylen	600
10	"	575
25	"	540
1 %	H_2S	373
8	"	304

2. Normala förbränningshastigheter.





Normala förbränningshastigheten är som synes högst i en blandning med luftunderskott, och den är lägre ju större kolväte-molekylen är.

Temperaturinflytandet på v_n är ej särdeles stort.

För koloxid/luft: vid 20 °C: 42 cm/sek, vid 460 °C: 85 cm/sek.
" metan/luft: — " — 31 — " — 430 °C: 64 — " —

förbränningen i ett rör, som upphettas, ökar gasvolymen och därmed gasens strömningshastighet fortare än förbränningshastigheten, dvs. lågan blåses ut ur röret. (Om detta inträffar även vid höga temperaturer är ej bekant.)

Tryckberoendet är ännu ringare.

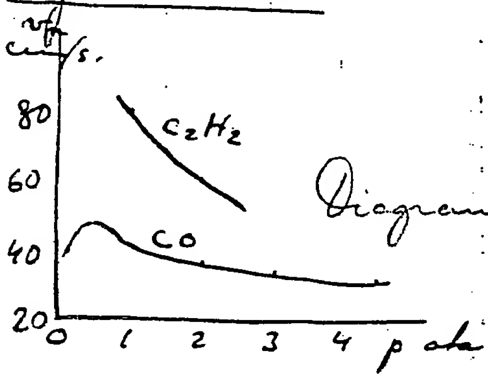
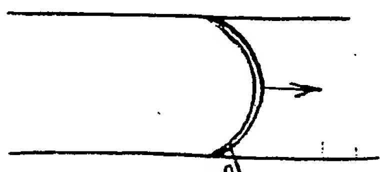


Diagram 3.

3. Förbränningshastigheten i rör.

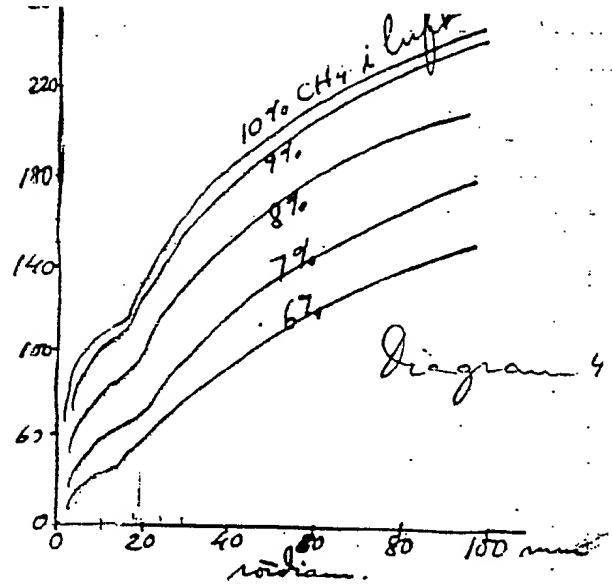


flamma

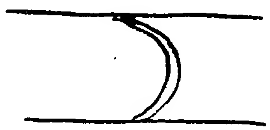
Flamman är alltid buktig framåt. Då den förbrända mängden är prop. mot flammans yta, blir den på tidsenhet förbrända volymen gas större än den i motsvarande del av röret befintliga gasvolymen, dvs. för att flamman skall vara stabil till sin form (men fortfarande röra sig framåt i röret), måste brännbar gasblandning strömma mot flamman hela tiden. Flammans fortplantningshastighet relativt gasen, dvs. förbränningshastigheten v_{ff}, blir sålunda större i ett rör än i öppen volym. Hur mycket större den blir beror bl.a. rördiametern, men den är i allmänhet ~~omkring~~ ungefär $2{,}0 \times v_n$. Detta gäller mitten av röret. Vid rörväggen kyles så mycket värme bort, att hastigheten nedsättes.

Rördiameterns inverkan framgår av följ. diagram:

mm, fortplantar sig flamman överhuvudtaget ej.



4. Instabilitet vid förbränning i rör.

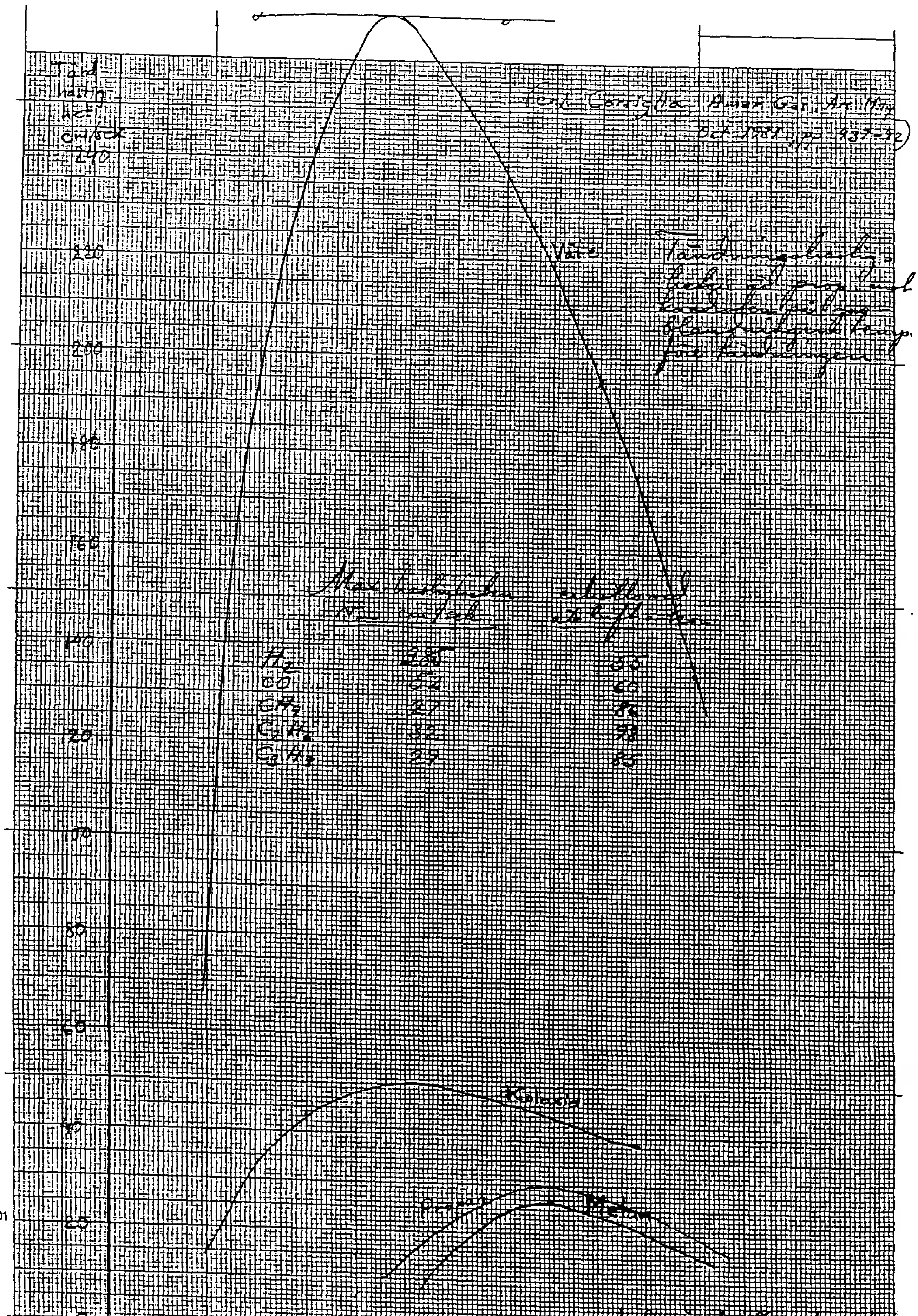


Den förbrända gasvolymen är, som sagts, prop. mot flam-ytan. Om denna störes, genom t.ex. en oregelbundenhet i gas-tillströmningen, förstoras flam-ytan, och mängden förbränd gas ökar, vilket verkar i samma riktning som störningen själv, dvs. störningen förstärkes, dvs. svängningar och till slut detonationer kan uppkomma. Smärre störningar dämpas dock bort av röret.

7. Om inverkan rördiametern:

Ju grövre rör, desto större effektiv förbränningshastighet, d.v.s. desto större gasmängd kan brännas per timme och cm^2 tvärsnitt. En 1" brännare har ett tvärsnitt av 4.9 cm^2 och en 3/4"-brännare ett tvärsnitt av 2.8 cm^2. Enl. diagr. 4 är v_{eff} för 10% CH_4 i luft i ett 1"-rör ca 148 cm/s och i ett 3/4"-rör ca 135 cm/sek, d.v.s. i 1"-röret förbrännes per sek. $148 \cdot 4.9 = \underline{\underline{722}}$ cm^3 gas och i 3/4"-röret $135 \cdot 2.8 = \underline{380}$ cm^3 gas, d.v.s. blott 53% av 1"-rörets kap.

Den angivna värmemängden är proportionell mot dessa volymer.



Tänd-
hastighet
cm/sek
Väte
Max hastighet
cm/sek
H2 285 55
CO 52 60
CH4 27 86
C2H4 32 78
C3H8 27 85
Koloxid
Propan
Metan
Luftmängd i % av teoretiska

Tändhastighet, cm/sek.

% syre	% gas i blandning med syre	väte	acetylen	stadsgas	propan
90	10	—	750		7: 300
80	20	—	1150		370
70	30	—	max. 27: 1340 1250		13: 385 270
60	40	—	900	45: 600	22: 200
50	50	600	740	670	
40	60	800	—	50: 710	
30	70	890	—	60: 660 65: 500	
20	80	800	—	70: 300	
10	90	350	—		
0	100	—	—		

% luft	% gas i blandning med luft	H_2	C_2H_2	Stadsgas	C_3H_8
95	5				2: 20
90	10				5: 30
85	15				6: 20
80	20	80		10	10: 10
75	25	150		50	
70	30	200		65	
65	35	~~[illegible]~~		68 60	
60	40	260		20	
55	45	265			
50	50	255			
45	~~[illegible]~~55				
40	60	180			
35	65				

Tändhastigheter i cm/sek

		~~[illegible]~~ % syre resp. luft av teoretisk mängd	Syrgas: Väte	Stadsgas	Propan	Luft: Väte	Stadsgas	Propan
	50							
	60							
	70							
	80							
	90							
stökiom. blandn.	100							
	110							
	120							
	130							
	140							
	150							

KALKYL FÖR LINS-METODEN.

A. Kostnaden för att tillföra berget en miljon BTU.

Värme tillföres berget genom förbränning av gas med luft i brännare, nedsatta i borrhål. Ett stort antal kombinationer av hålavstånd, brännareffekt och bränntid är tänkbara. I kalkylen nedan förutsättes att ungefär de förhållanden, som råder i Santa-Cruz-fältet tillämpas.

Priser och löner, gällande i Californien för närvarande, har använts. Räntan på investerat kapital antas vara 5 % per år och underhållskostnaderna för utrustning 4 % per år. Utrustningens livslängd är bedömd från fall till fall. Drifttiden per kalenderår antas bli 7900 timmar (90 % availability).

1. Borrhålet.

Borrning (inklusive rörsättning), 60 fot á 0,35 $	21.00 $/hål
Omborrning och uppdragning av ytterröret efter driftperiodens slut, 60 fot á 0,35 $	21.00 "
Cementering runt gasröret	2.00 "
Montagearbete (anslutning till ledningsnät för bränsle och pyrolysgas)	2.00 "
Andel i kostnad för termometerhål (ett dylikt behövs för 20 - 100 brännarhål)	1.00 "
Summa	48.00 $/hål

Antalet borrhål per acre beror på hålavståndet. Eftersom 57 · 10^{6} ^{9} BTU skall tillföras per acre (inklusive värmeförluster uppåt och nedåt), erhålles:

Hålavstånd, fot	8	10	15	20
hål per acre	790	500	223	126
borrhålskostnad, $/acre	38.000	24.000	10.700	6.000
" $/$10^6$ BTU	0,665	0,420	0,188	0,106

2. Rören.

Det har visat sig att rören kan upptagas och användas ånyo. 3 års genomsnittlig livslängd antages. Ytterröret antages vara av 5 % Cr, 0,5 % Mo, 1,5 % Si - kvalitet.

20 fot gasrör (oleg.) à 0.80 $	16.00 $/hål
60 fot ytterrör (leg.) à 2.50 $	150.00 "
Summa	166.00 $/hål

Per drifttimme antages brännaren kunna inmata 25.000 BTU, varför kostnaden blir (med ränta, underhåll och avskrivning) 0,0083 $/drifttimme = = 0,332 $/$10^6$ BTU.

3. Armatur, fasta ledningsnät m.m.

Andel i fasta rörnät för tillförsel av

bränsle och bortförsel av pyrolysprodukter		15.00 $/hål
kopplingar, ventiler etc.		5.00 "
	Summa	20.00 $/hål

För dessa poster räknas med 10 års avskrivningstid, varför kostnaden blir 0,017 $/10^6 BTU.

4. Brännaren.

Brännaren kostar, inklusive nedledningsrör och anslutningsdetaljer 52,00 $/st.

Den antas kunna användas i 3 år med en inmatning av 25.000 BTU/drifttimme, varför kostnaden blir 0,096 $/10^6 BTU.

5. Kompressorstationen.

En miljon BTU, tillfört tjärsandslagret, motsvarar ca 1,2 $\cdot$ 10^6 BTU i gasen eller 1330 cuft gas av värmevärdet 900 BTU/cuft (som gäller för såväl pyrolys- som naturgas). Motsvarande luftmängd är 12.000 cuft. Sammanlagt skall alltså 13.330 cuft gas + luft komprimeras till 12 psig (brännaren behöver 7 - 10 psig). Enligt kompressortillverkare kan man utan risk blanda gas och luft före kompressionen. En lämplig enhet skulle vara en kompressor med en kapacitet av ca 600 cuft/min, som räcker för 100 brännare à 25.000 BTU/h. En komplett enhet kostar:

kompressor		3.000 $
elmotor (30 hkr) + varvtalsvariator		1.000 $
blandningsregulator för gas - luft		700 $
el- och gasledningar, fundament, montage		300 $
	Summa	5.000 $

Denna enhet antas ha 10 års avskrivningstid, varför den fasta kostnaden blir 0,105 $/timme = 0,042 $/10^6 BTU.

6. Kompressordriften.

Effektförbrukningen för en kompressorstation för 100 brännare är ca 18,5 kW, som vid kraftpriset 1,0 cts per kWh motsvarar 0,185 $/drifttimme eller 0,074 $/10^6 BTU.

Kompressorstationen kan göras praktiskt taget helautomatisk. Den tillsyn, som behövs, inkluderas i Arbetslöner.

7. Löner och administration.

Arbetsstyrkan för en 1000-brännaranläggning uppskattas bli 2 dagtidsarbetare (för underhåll) och 1 man per skift (för kompressor-, brännar- och pumpövervakning) För borrning erforderlig personal är inkluderad i borrkostnaden.

arbetare, 40 timmar/dygn à 2,00 $	= 80,00 $/dygn
arbetsledare (eller driftingenjör)	= 20,00 "
administration, 20 % av lönekostnaden	= 20,00 "
Summa	120,00 $/dygn

Kostnaden blir alltså 0,200 $/10⁶ BTU.

Sammandrag

vid hålavståndet	kostnad i $ per 10^6 tillförda BTU 8 fot	10 fot	15 fot	20 fot
1. Borrhålet	0,665	0,420	0,188	0,106
2. Rören	0,332	0,332	0,332	0,332
3. Armatur, ledningsnät	0,017	0,017	0,017	0,017
4. Brännaren	0,096	0,096	0,096	0,096
5. Kompressorstationen	0,042	0,042	0,042	0,042
6. Kompressordriften	0,074	0,074	0,074	0,074
7. Löner och administration	0,200	0,200	0,200	0,200
Summa	1,426	1,181	0,949	0,867

Anmärkning.

Det har här antagits att fältet är självförsörjande med bränslegas. Om så ej blir fallet kan tillsatsbränsle (naturgas) köpas för 0,50 $/$10^6$ BTU.

B. Oljeutvinningen per tillförd miljon BTU.

För att upphetta 1 cuft tjärsand till pyrolystemperatur åtgår teoretiskt 21,000 BTU. Om oljeutbytet är 4 vikts-% blir utvinningen 0,71 barrel per tillförda 10^6 BTU och om oljeutbytet är 6%, erhålles 1,08 barrel per 10^6 BTU.

I Santa Cruz-fyndigheten är genomsnittliga tjärhalten 8 vikts-%, varav man kan vänta sig att utvinna mellan 50 och 65 % som olja. För säkerhets skull räknas här med den lägre siffran, d.v.s. med 4 vikts-% oljeutbyte.

I ett enhålsförsök är värmeförlusterna till omgivningen mycket stora. Det kan matematiskt visas att endast 1,25 % av det tillförda värmet användes för verklig pyrolys. Sålunda erhålles per 10^6 BTU blott 0,0089 barrel. I enhålsförsök L 3 erhölls ca 0,02 barrels per 10^6 BTU, men tjärsanden var där rikare. (Den del av borrkärnan, som kunde tillvaratagas, höll ca 9% tjära.

I ett sjuhålsförsök är förlusterna till att börja med lika stora som i sju separata enhålsförsök, men efterhand som brännarnas samverkan kommer till synes, sjunker förlusterna, relativt sett, till ett minimum av ungefär 60 % av det tillförda värmet. Per 10^6 BTU erhålles då ca 0,28 barrels olja.

Efter lång tid flyter de sju brännarnas verkningar ihop till ungefär samma resultat, som skulle erhållas med en enda, sju gånger större brännare. Förlusterna motsvarar då ånyo förhållandena i ett enhålsförsök.

I försök L 72, där genomsnittliga tjärhalten var relativt låg, 7,3 %, erhölls totalt 4,16 barrels olja per 191· 10^6 tillförda BTU eller 0,022 barrels/10^6 BTU. Korrektion till 8 % tjärhalt höjer siffran till 0,024 barrels/10^6 BTU.

I en mång-brännaranläggning beskriver de procentuella värmeförlusterna en liknande kurva som i en sjuhålsenhet med den skillnaden att minimiförlusten är konstant, så länge fältet kontinuerligt fortskrider framåt. Vid avslutning av ett begränsat fält stiger förlusterna åter.

För hundrahålsfältet L 8 har det beräknats att totalt 3400 barrels skulle erhållas med en inmatning av 11.900 · 10^6 BTU (fältets genomsnittliga tjärhalt = 7,3 %). Oljeutvinningen skulle sålunda bli 0,286 barrels/10^6 BTU. Under den tid fältet hade någotsånär konstanta driftförhållanden erhölls ca 0,09 barrels/10^6 BTU.

I en full-skala-anläggning med kontinuerlig fältflyttning beror förlusterna huvudsakligen på fältbredden och vandringshastigheten. I ett 2000 fot brett fält med 10 fots hålavstånd blir förlusterna ca 35 %, d.v.s. vid ett oljeutbyte av 4 vikts-% erhålles 0,46 barrels/10^6 BTU.

C. Sammanfattning.

De ovan gjorda kalkylerna visar sålunda att vid en fullstor anläggning med 10 fots hålavstånd tillverkningskostnaden för 0,46 barrels olja blir 1,18 $, eller för 1 barrel 2,55 $. Därtill skall läggas kostnaden för kondensering och lagring, som i en stor anläggning är blygsam, säg 5 cts/barrel.

Oljan skulle alltså kosta, fritt anläggningen 2,60 $/bbl.

För den olja, som hittills sålts, har erhållits 3,11 $/bbl. Den har emellertid varit något tyngre (spec.vikt 0,904) än vad som kan väntas från en fullstor anläggning (spec.vikt ca 0,880), varför försäljningspriset torde bli något högre. Transporten till kunden (raffinaderiet) kan väntas kosta max. ca 10 cts/barrel.

Kostnaden för gasens svavelrening har ej inkluderats i kalkylen, då den bör kunna bäras av det utvunna svavlet, för vilket ingen kreditering gjorts. Per m^3 olja blir svavelproduktionen av storleksordningen 30 kg.

Närkes Kvarntorp den 4 maj 1957

Gösta Salomonsson

Överingenjör

and specific heat. As good determinations are reported in the literature, no accurate measurements were made. The reported data are:

specific heat	0,22 cal/g, °C
heat conductivity	0,0035 cal/cm, °C, sec.
specific gravity	2,0 g/cm³

Measurements in connection with the LINS model tests /8 and 9 above/ were in agreement with what could be calculated from these data.

11. Preliminary calculations for a LINS-field.

From the data and observations obtained in the above-mentioned tests, some fundamental calculations could be made, a summary of which is given below.

As the oil yield is of utmost importance a comparison is made between different oil recovery methods. The figures for the methods, numbered 1) to 5) are given by Blair in his official report and are results of semi-commercial tests. The figures for the LINS Method are obtained in a small-scale tests and are thus not fully comparable, which must be remembered in the further calculations.

No.	Process sequence	In-put tar as tar sand	Out-put of liquid products gas oil	gasoline	butane
1	hot-water-sep.+dehydratation+ +conventional coking	100 bbls	48 bbls	17 bbls	1 bbls
2	hot-water-sep.+fluid-bed coking	100	67	7	(x
3	cold-water-sep.+dehydratation+ +conventional coking	100	57	17	2
4	hot-water-sep.+fluid-bed catalytic coking	100	8÷30	7-22	(x
5	fluid-bed coking tar sand	100	79	6	1
6	the LINS Method	100	52	28	(x

(x not determined.

As far as yields concern the LINS Method thus is promising.

Another important factor is the heat balance for the process. From the specific heat, specific gravity and temperature for complete pyrolysis (750°F), obtained in the pyrolysis test, combined with some reaction-kinetic studies) it can be calculated that the theoretical heat need will be about 180 BTU/lb of tar sand. From the experiences in the Ljungstrom field at Kvarntorp it can be learnt that the actual need will be a little higher, say 220 BTU/lb, because of heat losses to the surroundings etc. (These losses are of course relatively smaller the larger the field area is.) The heat content of the uncondensable gas, liberated from 1 lb of tar sand is about 350 - 400 BTUs.

In a gas-fired element tube an overall calorific efficiency of 70–75 % may be obtained, resulting in between 250 and 300 BTUs available in the rock layers for heating. With the above mentioned gas yield there will thus be enough heat available to make the plant selfsupporting. Even if the gas yields in some districts would be 10–15 % smaller, which cannot be predicted, there is a margin between fuel production and consumption. It may thus be concluded that gas-fired elements would be the best alternative for heating the tar sand. Also the coke combustion method may be possible, but of these two the former has the advantage of not effecting at all the quantity or quality of the recovered oil, which the coke combustion method may do to some extent. Of course, if also the gas has a high market value, the coke should be utilized in the above-mentioned manner.

<u>Hole pattern for the heating element. Heat distribution.</u>

The most conveniant way to apply the heat is in the shape of vertical element tubes, inserted in drillholes, spread over the field. (There are methods of supplying heat equally over a large, horizontal surface, but these methods are not suitable for a mineral, like the tar sand, which has no horizontal lamination.) The drill-holes should be equally distributed over the whole surface. In fig. 4 are shown some different drill-patterns, which could be used. In the Ljüngstrom field at Kvarntorp the hexagonal pattern is used. The heating elements are arranged in the corners of each hexagone and the gas outlets are drilled in the centres. For reasons, which are mentioned below, it seems possible to combine heating element and gas outlet in one unit in the tar sand. In that case the triangular pattern will be the most suitable one.

The heat distribution is slow as well in tar sand as in shale. The reported heat conductivity of tar sand, 0,0035 c.g.s. units, is exactly the same as for the Kvarntorp shale (in horizontal direction). Thus the heat distribution around the heating elements will be the same in both cases. Exact heat transfer calculations are made in Appendix 1 to this report. From this it can be found that if a triangular hole pattern with 10 ft hole distance is used and a heating effect, corresponding to about 1100 BTU/hour and foot of element length, is supplied, the required temperature, 750°F, is reached in every point of the sand after about 2400 hours heating (14 weeks). Each element has to heat about 43 sq.ft of the field area. If drilling or tubing costs are high it might be more economical to have a sparser hole pattern and correspondingly larger heating period for each element. Heating periods of up to one year may not

be considered as in any way abnormal. Also the heating effect supplied to each foot of the element may be changed in order to get an optimum combination of drilling costs, element costs, fuel consumption, supervision labor, etc. (It may be worth mentioning, that as well hole distance as heating effect in the Ljungstrom field at Kvarntorp have been successively changed in direction towards lower overall costs.)

Gas outlets.

In the shale field the horizontal laminations in the shale layers open up during the heating and thus offer suitable flow paths for the vapours and gases. On the other hand the flow in vertical directions is more or less restricted. The gas outlets are therefore drilled through the whole shale layer. In the tar sand there exists no lamination and the collection of oil vapours and gases thus offers a special problem. The tar sand is in itself impermeable but by heating the tar becomes less viscous and starts to flow if subjected to gas pressure. When sand is heated to pyrolysis temperature the evolved vapours act in three ways to facilitate their flow:

a) they create a superpressure in the neighbourhood of the element.
b) they transfer heat to the tar in the surroundings
c) they condense partly in the colder parts of the rock and the condensate acts as a solvent for the tar, forming a less viscous solution.

As far as smaller model tests have shown it is well possible to take out the vapours from any desired point in the field. It is of course not possible to state without experiments in the actual field, that this conclusion is valid also for larger distances between the element holes. It is always possible, however, to collect the vapours at the point, were they are liberated in the rock. As the coke left behind is highly permeable for gases it is also possible to arrange the gas holes in such a manner that there is an unbroken flow path through "coked" parts of the sand to the holes. This is the case if the gas outlets are arranged concentrically round the element tubes.

Summary:

In all respects that have been possible to investigate on a laboratory scale and in small model tests the LINS Method for oil recovery from tar sands. The process will be thermally self-supporting and good yields are obtained. The high gasoline content of the obtained oil is remarkable.

Sammandrag av försök med olika brännare i hål L 22

För-sök nr	Brän-nar-typ	Brännarrör diam. tum	Brännarrör längd fot	Strålningsskydd 1 diam. tum	Strålningsskydd 1 längd fot	Strålningsskydd 2 diam. tum	Strålningsskydd 2 längd fot	Rökgascirk.% genom ejektor	Rökgascirk.% mellan brännarrör o.stråln.-skydd	Avgivet värme BTU/h	Temp. efter 3 tim. °F mitt för konan t_0	Temp. efter 3 tim. °F 8 fot nedom konan t_8	Temp. efter 3 tim. °F 16 fot nedom konan t_{16}	Temp.ökningskvoten $\frac{t_0 - 70^0}{t_{16} - 70^0}$	Temp.ökningskvoten $\frac{t_8 - 70^0}{t_{16} - 70^0}$
1	A	1	23,5	-	-	-	-	-	-	22.500	900	275	110	21	5,1
2	B	1	23,5	-	-	-	-	15	-	22.500	360	270	120	16	4,0
3	C	1	23,5	1,75	10	-	-	-	-	22.500	900	310	175	8	2,3
4	C	1	23,5	1,75	10	-	-	15	-	22.500	810	310	150	9	3,0
5	C	1	23,5	1,75	10	-	-	15	-	18.500	755	230	115	15	3,6
6	C	1	23,5	2,00	10	-	-	15	-	22.500	850	265	145	10	2,6
7	C	1	23,5	2,25	10	-	-	15	-	22,500	775	280	150	9	2,6
8	D	1	23,5	1,75	10	2,00	4	15	-	22.500	755	280	170	7	2,1
9	E	1	8	1,75	24	-	-	25	1)	22.500	880	380	130	14	5,2
10	E	1	8	1,75	21,5	-	-	25	1)	20.000	785	355	130	12	4,8
11	E	1	10	1,75	21,5	-	-	15	1)	20.000	770	275	140	10	2,9
12	E	1	10	1,75	21,5	-	-	25	1)	35.000	940	400	235	5	2,2
13	E	1	10	1,75	21,5	-	-	15	2)	20.000	925	275	145	11	2,7
14	E	1	10	1,75	21,5	-	-	25	2)	35.000	1010	410	230	6	2,1
15	E	1	10	1,75	21,5	-	-	25	3)	20.000	920	275	130	14	3,4
16	E	1	10	1,75	21,5	-	-	25	4)	20.000	935	280	140	12	3,0
17	E	1	15	1,75	21,5	-	-	25	1)	20.000	765	295	205	5	1,7
18	E	1	15	1,75	21,5	-	-	15	1)	35.000	(1000	400	310	4	1,4
19	E	1	8	1,75	15	-	-	25	1)	20.000	705	335	285	3	1,2

1) möjligen någon rökgascirkulation.
2) rökgascirkulationen ökad genom strypning av arean vid brännarrörets nedre ända.
3) rökgascirkulationen mellan rören omvänd (=uppåtriktad) p.g.a. injektorns placering.
4) sannolikt ingen rökgascirkulation.

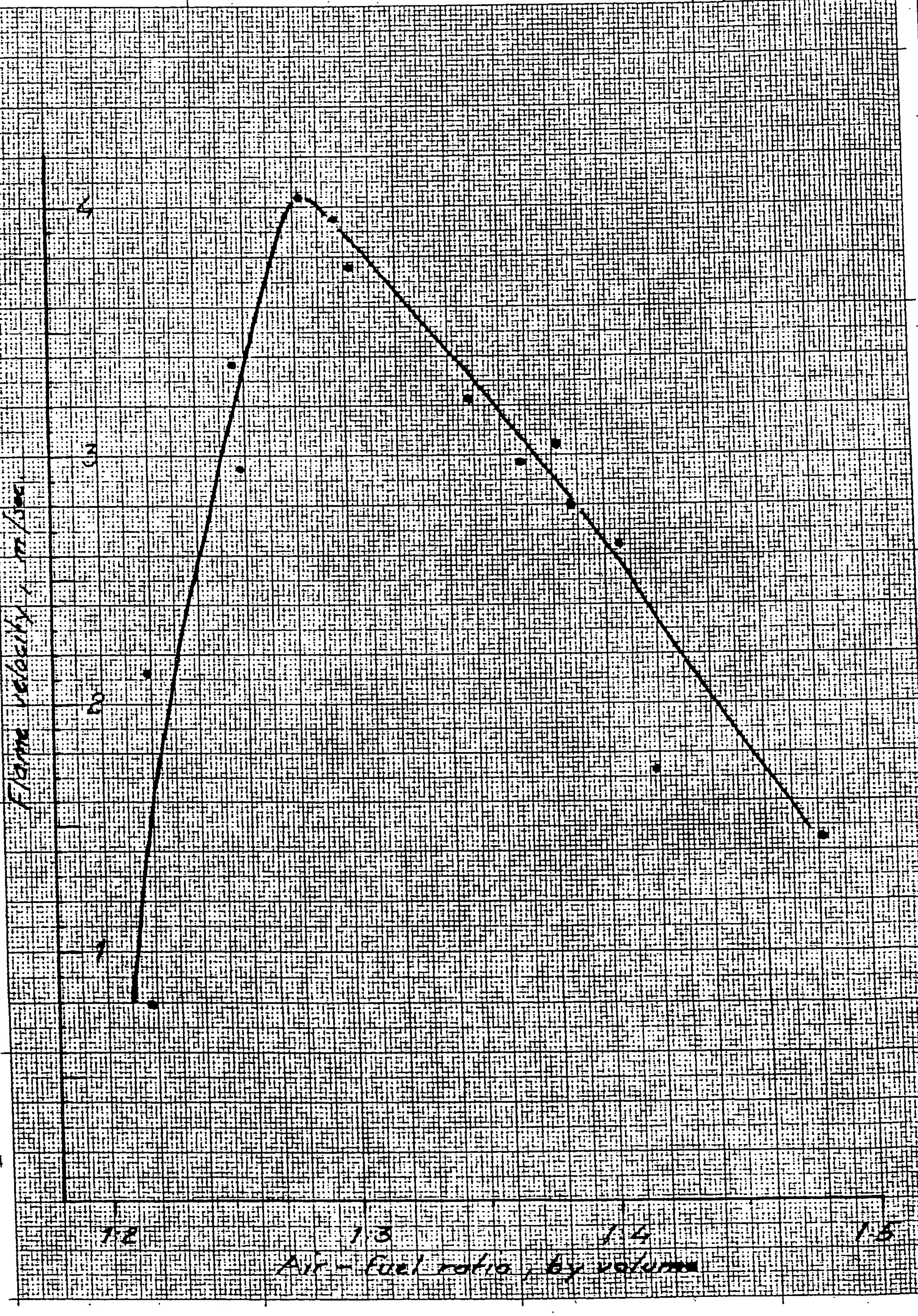
Förbränningshastigheter

Rörets diam.= 26,75 mm.
Gränsvärden där lågans hast.= gashastigheten

gasavl.	korr	luft	förh.	gas+luft	v m/sek
300	340	800	1:2,35	1140	0,59
1000	1100	2600	1:2,36	3700	1,91
1300	1420	3900	1:2,75	5320	2,74
1500	1650	4500	1:2,73	6150	3,17
1700	1850	5600	1:3	7450	3,84
1600	1750	5500	1:3,14	7250	3,74
1500	1650	5250	1:3,20	6890	3,55
1150	1260	4600	1:3,65	5860	3,02
1000	1105	4250	1:3,85	5355	2,76
1000	1100	4400	1:4	5500	2,84
900	1000	4050	1:4,05	5050	2,60
800	900	3800	1:4,22	4700	2,42
500	550	2400	1:4,37	2950	1,52
300	340	1700	1:5	2040	1,24

tube, at 10°C and 1 at.

4·1·1954.NF



514 A4
1 mm
ESSELTE
4441

Beträffande gasbalansen vid en LINS - anläggning

I hundrahålsförsöket i Santa Cruz inmatades $19.500 \cdot 10^6$ BTU och producerades $4.429 \cdot 10^3$ kubikfot gas. Efter uppvärmningens avbrytände har ytterligare $50 \cdot 10^3$ kubikfot gas erhållits. Genom markläckage och vid störningar i apparaturen har uppskattningsvis ytterligare $500 \cdot 10^3$ kubikfot gas bortgått, varför ungefär $5.000 \cdot 10^3$ kubikfot gas torde ha producerats. Häri är fortfarande ej inräknat eventuella förluster horisontellt ut i omgivande tjärsandslager.

Gasens värmevärde är enligt stickprovsanalyser omkring 1.000 BTU/kubikfot. Följaktligen skulle gasproduktionen motsvara ca $\underline{5.000 \cdot 10^6 \text{ BTU}}$ eller ca 25 % av bränsleförbrukningen.

Enligt erfarenheter från Ljungströmsanläggningen i Kvarntorp är energiförbrukningen vid ett fält i 100-hålsskala ca 14 kWh/liter olja, att jämföra med 6,5 kWh/liter olja i den nuvarande 2.400-hålsanläggningen. Minskningen sammanhänger med minskningen av förluster till omgivningen. Om samma proportioner antagas gälla i tjärsand, vilket rimligen bör vara fallet, skulle bränsleförbrukningen i ett stort fält för samma kvantitet produkter, som erhölls i 100-hålsförsöket, bli $\frac{19.500 \cdot 10^6 \cdot 6,5}{14} = \underline{\underline{9.000 \cdot 10^6 \text{ BTU}}}$.

Produktionen av $5.000 \cdot 10^6$ BTU skulle då täcka ca 55 % av bränslebehovet.

Tjärhalten i försöksfältet i Santa Cruz är i genomsnitt 8 %. Gasutbytet stiger i direkt proportion till tjärhalten. Följaktligen skulle en 15 %-ig tjärsand ge $\frac{15}{8} \cdot 5.000 = 9.500 \cdot 10^6$ BTU gas, d v s mer än vad som behöves för att göra ett fält självförsörjande.

Närkes Kvarntorp den 2 mars 1959

Gösta Salomonsson

ÖK

Tar sand samples containing 12.56 % b.w. tar, were heated in a Fisher retort at a rate of 14.3°F/minute to the following temperature levels: 500°, 600°, 700°, 800°, 850° and kept at these for 2 hours. The products were collected and measured. Thereafter the solid residue from each test was assayed according to Fisher.

<u>Results:</u>

~~From~~ 100 gram (dry) sample

preheated to	500°	600°	700°	800°	850° F
yielded: oil, grams	0,00	0,32	1,71	7.50	8.25
gas, "	0.01	0.01	0.23	0.62	0.75
water "	0,00	0.00	0,00	0.08	0.23
residue "	99.17	99.61	97.52	91.95	90.69
<u>Fisher assay products from these residues:</u>					
oil "	8.00	8.11	6.13	0.70	0.00
gas "	0.82	0.85	0.77	0.39	0.25
water "	0.23	0.13	0.00	0.00	0.04
coke "	90.03	90.57	90.67	91.13	90.47

<u>Thus, overall yields</u>

oil grams	8.00	8.43	7.84	8.20	8.25
gas, "	0.83	0.86	1.00	1.01	1.00
water "	0.23	0.13	0.00	0.08	0.27
~~coke~~ "	~~90.03~~				
total volatiles, g.	9.06	9.42	8.84 ?	9.29	9.52

<u>Conclusions:</u> Within experimental accuracy:

1) preheating does not influence overall oil yields.
2) — " — might increase gas yields
3) tar decomposition is slow below 700°F.
After 2 hours heating at 600° 700° 800° 850°F
there still remains ~93% ~70% ~5% 0%
of the original tar.

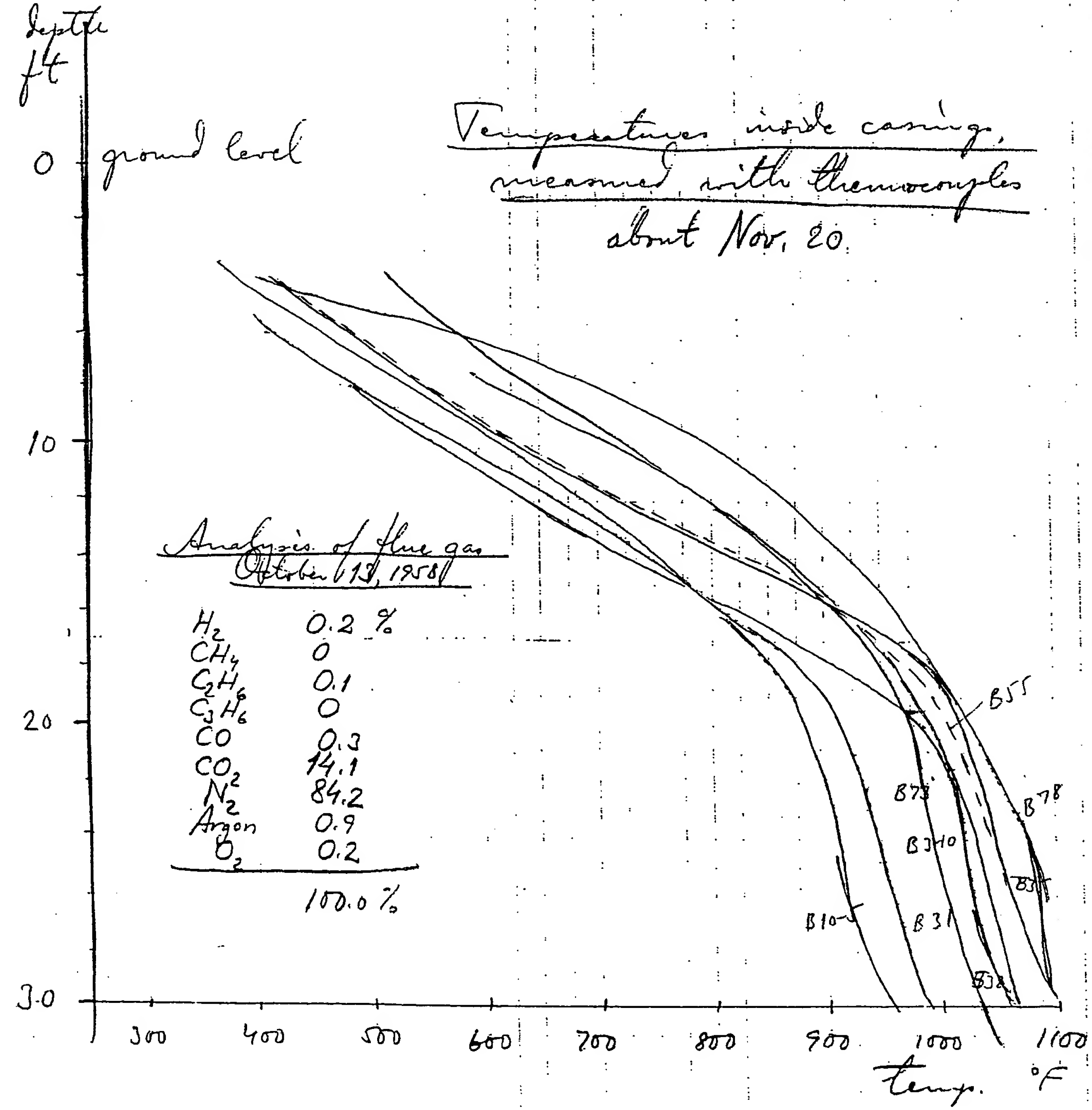
May 7, 1959
Sol

1. Temperatures of exhaust gas, measured inside the burner casings at ground level.

October 15-16 1958: average temp. = 322 °F
(highest 484°, lowest 204°F)

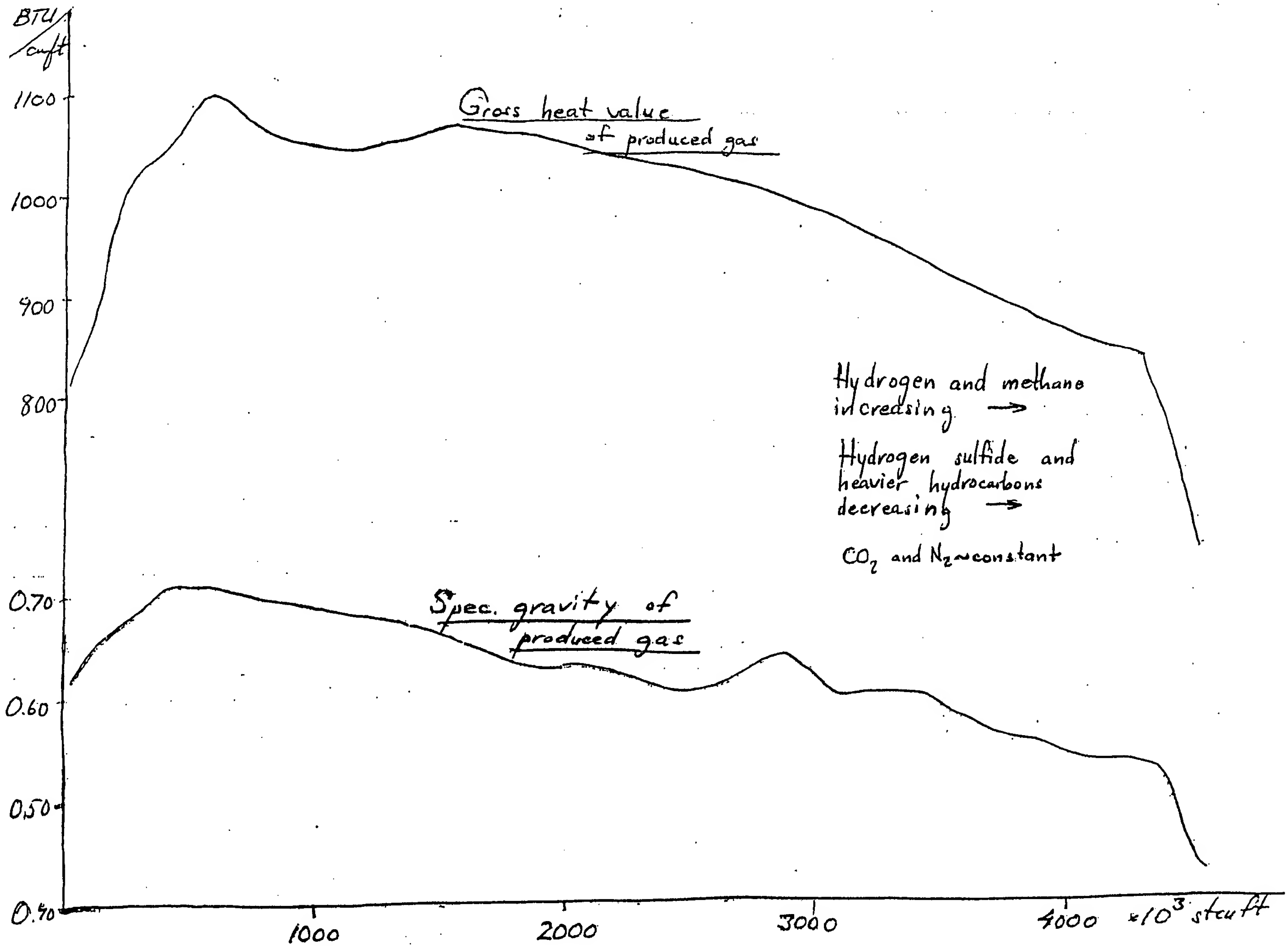
November 13-16, 1958: average temp. = 304 °F
(highest 475°; lowest 160°F).

Temperatures inside casing, measured with thermocouples about Nov. 20.



Analysis of flue gas
October 13, 1958

H_2	0.2 %
CH_4	0
C_2H_6	0.1
C_3H_6	0
CO	0.3
CO_2	14.1
N_2	84.2
Argon	0.9
O_2	0.2
	100.0 %



BTU/cuft
1100
1000
900
800
Gross heat value of produced gas
Hydrogen and methane increasing →
Hydrogen sulfide and heavier hydrocarbons decreasing →
CO2 and N2 ~ constant
0.70
0.60
0.50
0.40
Spec. gravity of produced gas
1000
2000
3000
4000
×10³ stcuft

Oil	2665 bbls =	415 tons =	15,880·10 BTU
Gas	4520 stcuft =	112 tons =	4,470 $\cdot 10^6$ BTU
Water	9232 bbls =	1615 tons =	—
Total		2142 tons	20,350 $\cdot 10^6$ BTU

Gas/oil ratio 1700 stcuft/bbl = 0.27 lb/lb
Water/oil ratio 3.46 bbl/bbl = 3.89 lb/lb

Heat output	20,350 $\cdot 10^6$ BTU
Heat input	19,550 $\cdot 10^6$ BTU
Net	800 $\cdot 10^6$ BTU

Gas analyses (composite)

H_2	39.6%	
H_2S	9.2	12.2
CO_2	2.3	
N_2+CO	0.7	
CH_4	28.6	42.9
C_2H_6	6.7	
C_3H_8	3.9	
iC_4H_{10}	1.2	
nC_4H_{10}	1.3	
iC_5H_{12}	0.3	
nC_5H_{12}	0.9	
C_2H_4	0.9	4.4
C_3H_6	0.6	
C_4H_8	1.7	
C_5H_{10}	1.2	
C_3H_4	0.1	0.2
C_4H_6	0.1	
Avg. C_6	0.7	
	100.0	

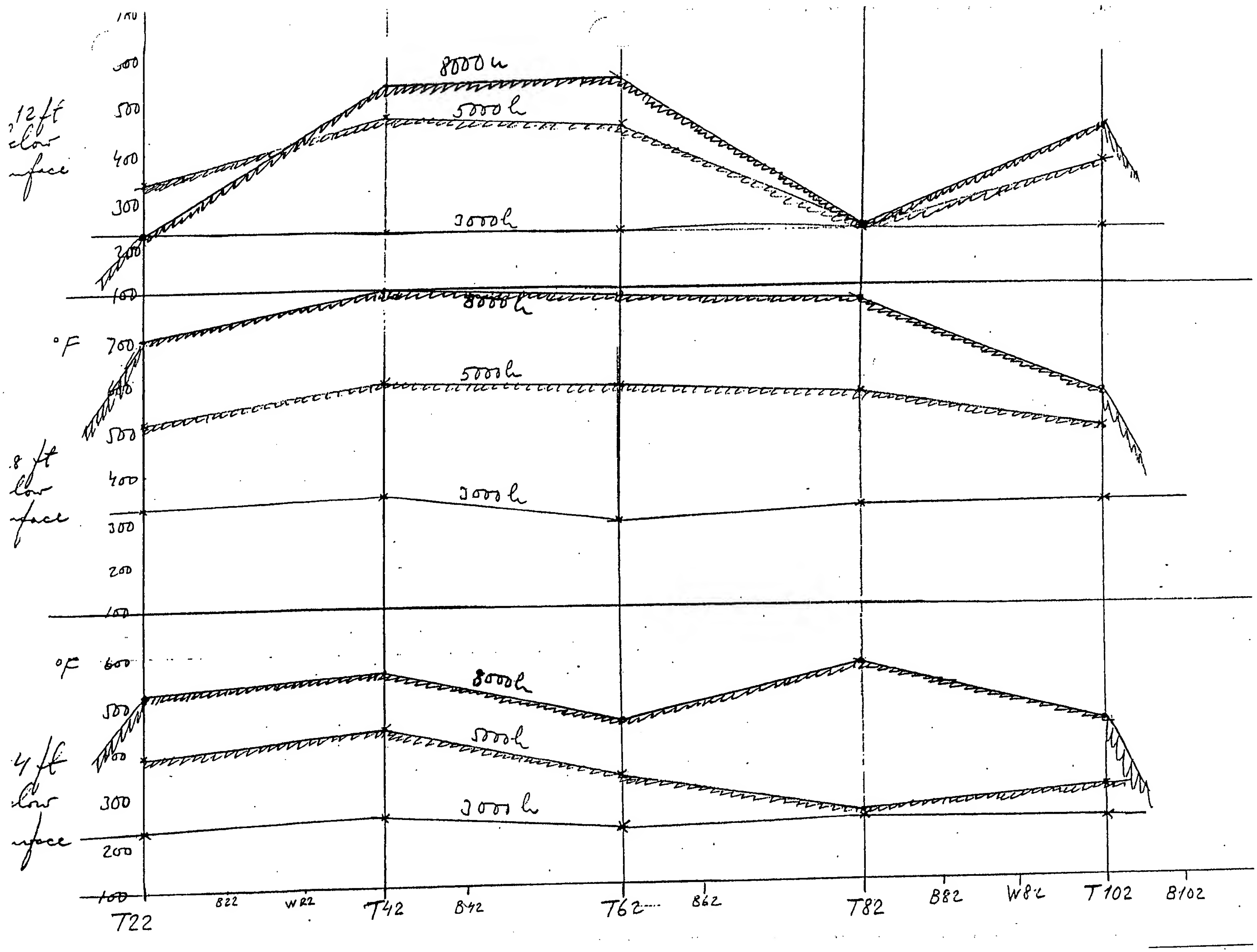
Calc. heat of combustion (gross) = 990 BTU/stcuft

Spec. gravity 0.65 (air = 1) = 0.840 kg/m^3 at 32°F

True spec. heat 0.024 BTU/stcuft at 65°F

Oil analyses

	July 26 Row 1	July 26 Row 7&8	July 26 Row 10	July 26 Total	Nov. 19 Total
Gravity °API	27.5	23.7	21.4	25.8	27.9
S, wt %	2.24	2.49	2.20	2.38	2.55
N, wt %	0.44	0.37	0.31	0.38	0.42
Dist. °F					
IBP	118	136	223	101	149
5%	250	260	355	270	240
10	320	345	425	320	290
20	415	450	500	400	385
30	465	575	550	470	460
40	500	565	585	520	575
50	540	605	630	570	560
60	590	650	680	615	600
70	645	705	750	670	635
80	720	775	835	740	680
90	815	855	930	840	735
95	900	920	—	—	~~[illegible]~~
Max.	915	970	965	910	755
Vol. rec.	95.5	96.0	92.0	94.0	93.5



12 ft below surface
8 ft below surface
4 ft below surface
°F
8000 h
5000 h
3000 h
T22
B22
W22
T42
B42
T62
B62
T82
B82
W82
T102
B102

(Temperatures measured in center of triangles between three adjacent burners.)

Temperature hole no	12 ft below surface			28 ft below surface			44 ft below surface		
	3000 h	5000 h	8000 h	3000 h	5000 h	8000 h	3000 h	5000 h	8000 h
T22	225	340	220	325	575	700	225	380	575
T24	230	360	415	335	530	740	230	360	55[illegible]
T28	285	420	455	400	565	715	265	445	540
T42	225	470	540	350	600	800	250	440	56[illegible]
T44	250	435	560	355	615	840	265	400	(~625 425
T48	300	430	490	385	640	810	265	410	(~522 270
T61	230	280	~350 220(*	310	475	630	220	270	~425 375
T62	225	450	530	290	585	790	225	330	500
T64	225	415	520	275	530	765	185	195	~570 375
T68	290	~500	525	330	530	775	230	325	~600 500
(T71	—	—	220	—	—	210	—	—	16[illegible]
T82	225	225	225	320	570	770	230	235	57[illegible]
T88	225	415	600	245	530	800	225	235	~60[illegible]
T102	225	365	440	325	490	~570 500*	220	295	44[illegible]
T108	226	226	410	225	220	545	220	220	~45[illegible]

*) Temperature dropped shortly before the end of the test probably due to water damage. ~ = extrapolated temperatures, assumed to correspond to undamaged reading

May 1, 19[illegible]
Sal

1. <u>Casings</u> (ASTM A213 - 53T Grade T5B from Babcock & Wilcox)

Ultimate strength 79,000 - 82,000 psi

Yield point 53,000 - 55,600 psi

% elong. in 2" 50-53

Hardness BHN 148-161

Chemical analysis:

C = 0.10 %

Mn = 0.42-0.44 %

S = 0.010-0.013 %

P = 0.018 %

Si = 1.39-1.49 %

Cr = 4.68-4.72 %

Mo = 0.45 - 0.54 %

Dimension: Schedule 40 (o.d. 2.875", wall thickness 0.203")

May 7 1959
Sal

Hours from start	Burner no.		Remarks
655	55	✓	Burner burned off by out-of-place flame.
1013	10-2		Condensing water clogged sand and cone burned off (& casing)
2470	89	✓	Broken supply tube
3173	72		Cone off
4135	82	✓	Flame in supply tube
4418	57	✓	— " —
4875	34	✓	— " —
5854	95	✓	Supply tube broken
6280	54	✓	Flame in supply tube. Cas. dam
6743	87	✓	Eroded hole in supply tube
6876	86		Cone burned off 3" from its top
6908	19	✓	Eroded hole in supply tube
7104	93	✓	Supply tube burned off
7198	98	✓	— " —
7389	73		Supply tube and casing damaged when coke in gas casing was burnt out
8021	49		Casing burned after power failure
After finished test	11	✓	Supply tube broken
— " —	28	✓	— " —

Power failures	3720	burnerhours	= 0,49 %
Instrument failures (prog.)	2040	"	0.26 "
Maintenance on fuel system	2160	"	0.28 "
Explosions in fuel lines	400	"	0.05 "
x) Burner failures	6413	"	0.83 "
Unplugging of coc gas wells	338	"	0.04 "
Repair of ground leaks	3842	"	0.50 "
" tankbushing leaks	471	"	0.06 "
Preventive burner mainten.	1825	"	0.24 "
xx) Unknown reasons	1572	"	0.20 "
Total off-time	22,787	burnerhours	= 2,95 % of total operating time.

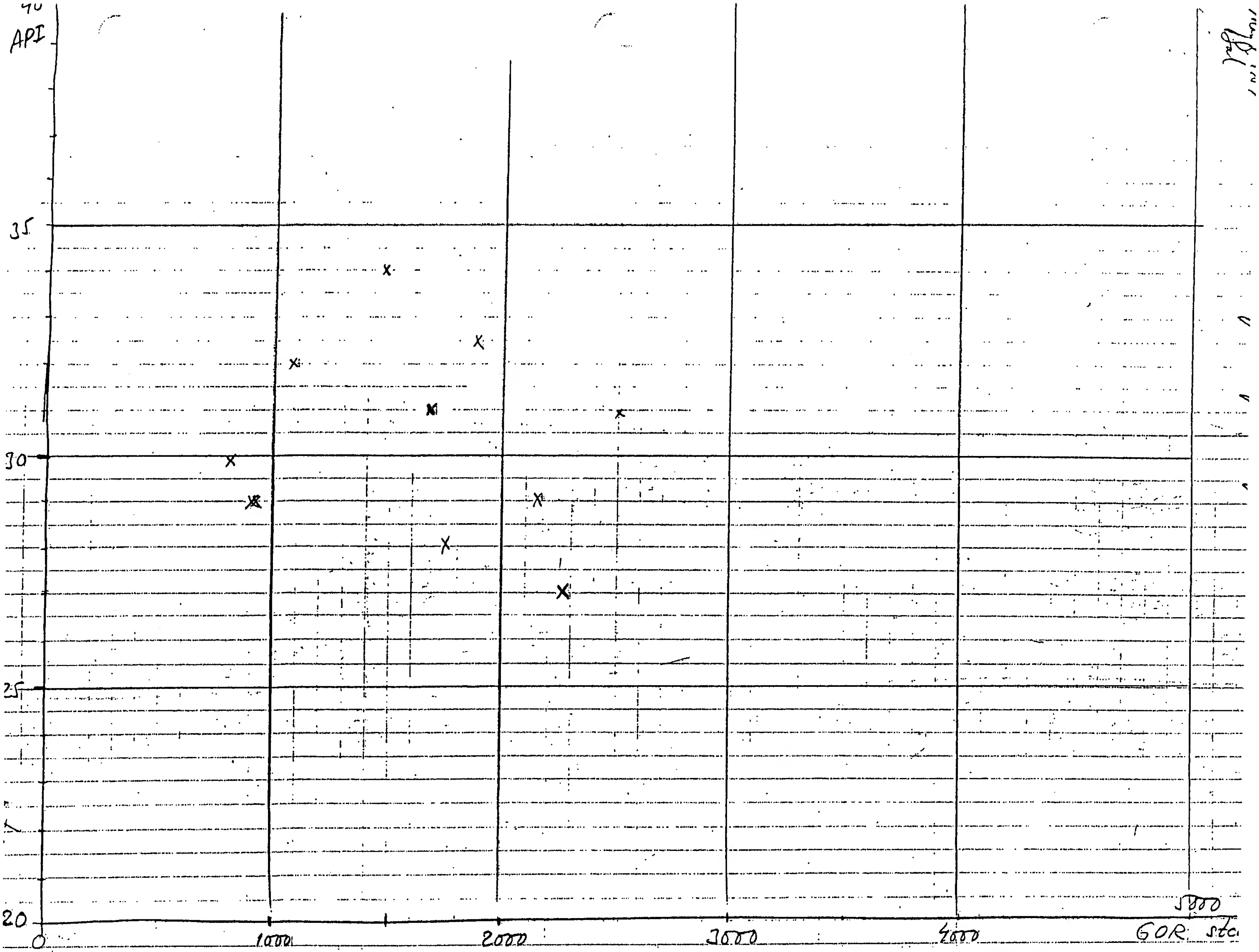
x) including: casings, cones, supply tubes, lighting difficulties

xx) Mainly during the first two test months. In many cases this was probably due to condensing water in the fuel lines.

May 2 1957
Bal

Date	Well	Gas/oil ratio	Water/oil ratio	Oil gravity °API
10.9.	B 53	6400	3.2	39
10.13.	B 53	6100	0.5	42
10.8.	B 73	1660	0.7	31
10.16.	B 4-10	2130	4.8	27
10.16.	B 8-10	2520	34	–
10.24.	B 1-10, 2-9, 10	910	0.2	29
11.1.	B 11, 12	1480	1.7	34
11.24.	B 12, 22, 23, 25	810	0.3	30
11.1.	B 14, 15, 17, 23, 25	1072	0.3	32
11.20.	~~B~~ Row 1, 2	2500	2.5	31
11.5.	B 23, 25, 27, 29	1890	0.7	33
11.4.	G 22, 24, 26, 28	1755	0.4	27
11.20.	— " —	2140	0.5	29
10.24.	B 1-10, 2-9, 10	910	0.2	29
11.24.	B 27, 28, 29	1570	0.3	27
11.13.	B 39, 3-10, 49, 4-10	1950	3.8	27
11.24.	G 41, 43, 49	1950	0.8	35
11.18.	B 57, 62, 63	1050	1.0	30
11.18.	B 55, 64, 65, 67	1350	1.3	26
11.18.	B 58, 59, 5-10, 69, 6-10	1260	0.8	30
11.5.	G 61, 63, 65, 67, 69	3160	1.0	39
11.24.	— " —	9550	1.6	41
11.18.	G 81, 83, 85, 87, 89	7300	0.4	27
11.21.	G 83, 87	5090	0.3	32
11.24.	B 95, 96, 97, 10-5, 6	1055	2.0	26
11.13.	B 9-10, 10-9, 10-10	2040	11.3	27
11.21.	G 92, 94, 96, 98	5000	2.3	29
11.24.	W 22	1330	0.3	30
11.26.	W 28	2440	0.7	32
11.25.	W 82	1960	3.4	28
11.26.	W 99	1900	6.8	24

etc.



40
API
35
30
25
20
0
1000
2000
3000
4000
5000
GOR scf

I

GOR 0-1000

910 —	28.6
810	29.8
910	28.6
470	23.0
474	27.4
943	26.0
(7) 782	22.7
5299	1881
ave. 760 —	25.8

GOR 2000-3000

2130 —	26.7
~~[illegible]~~	
2500	30.9
2140	28.6
2040	27.4
2440	31.5
2960	32.3
2650	29.2
2720	29.0
2080	28.9
2840	28.9
2800	36.0
(12) 2380	28.6
29680	3580
ave. 2470 —	30.0

GOR 3000-4000

3160 —	39.3
3070	27.0
3760	27.1
3540	27.0
3020	31.2
3900	33.3
3760	25.2
(8) 3100	27.0
27310	2371
ave. 3410 —	29.8

GOR 4000-5000

4290 —	28.5
4410	37.0
4040	30.8
4350	35.7
4210	24.1
4620	28.0
4680	29.7
4530	27.0
(9) 4350	23.1
39480	2639
ave. 4380 —	29.2

II

GOR 1000-2000

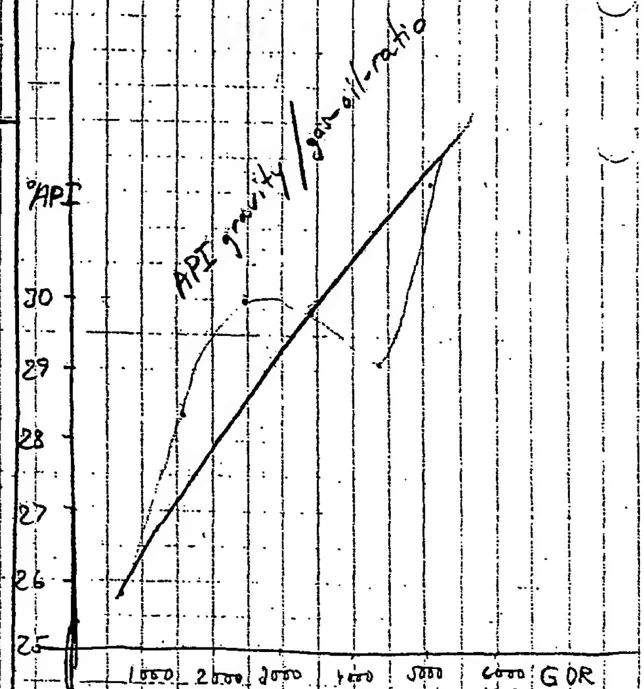
1660 —	31.3
1480 —	34.1
1072 —	31.9
1890	32.5
1755	26.9
1570	27.3
1950	27.0
1950	35.1
1050	29.7
1350	26.1
1260	29.6
1055	26.3
1330	29.9
1960	27.7
1900	24.4
1960	29.1
1840	31.2
1575	28.0
1330	25.2
1565	26.0
1305	25.7
1650	29.9
1720	27.9
1342	26.6
1525	26.0
(26) 1360	21.5
40344	7369
ave. 1550 —	28.4

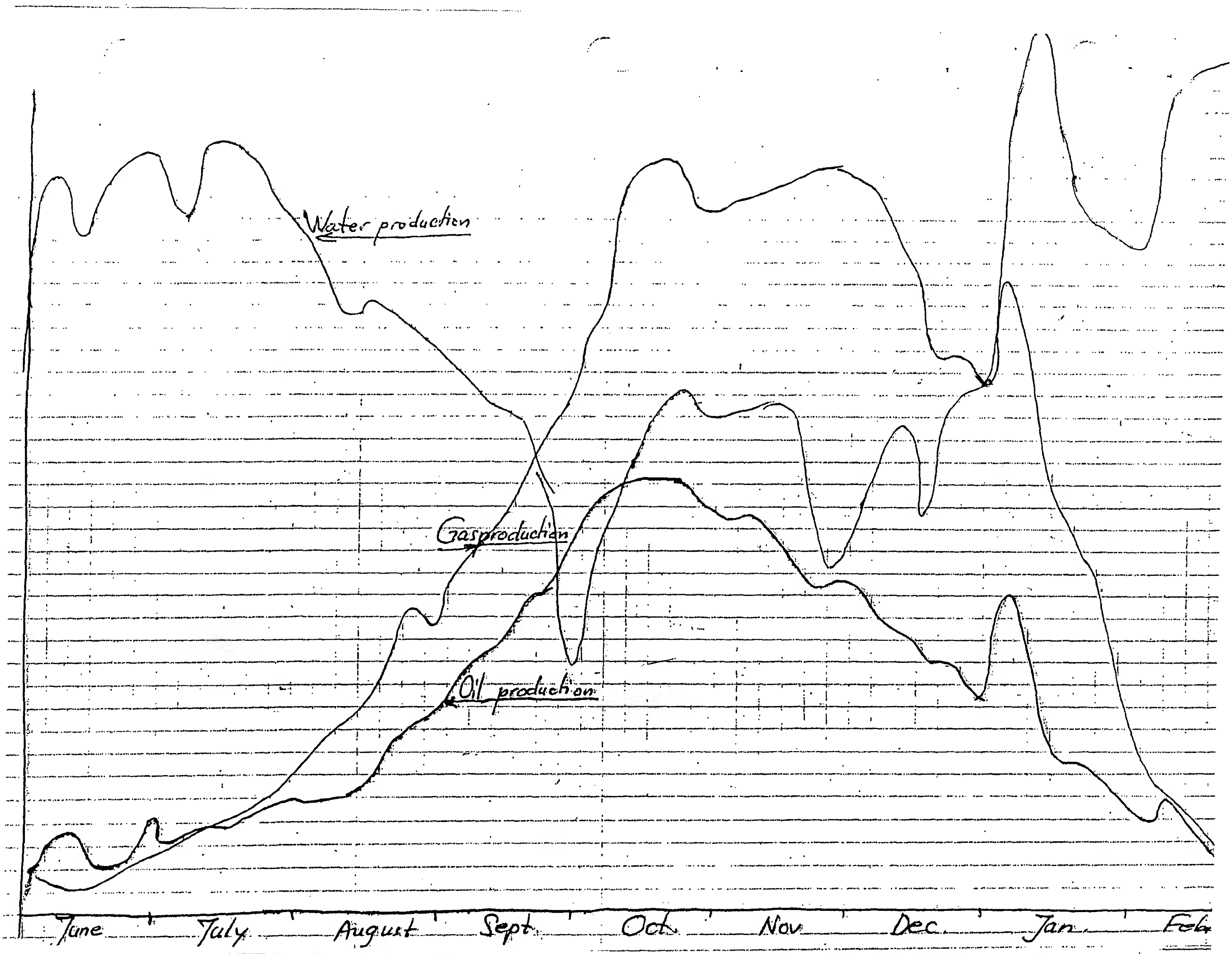
GOR 5000-6000

5090 —	31.5
5000 —	29.1
(3) 5050 —	34.6
15140	95.2
ave. 15050 —	31.7

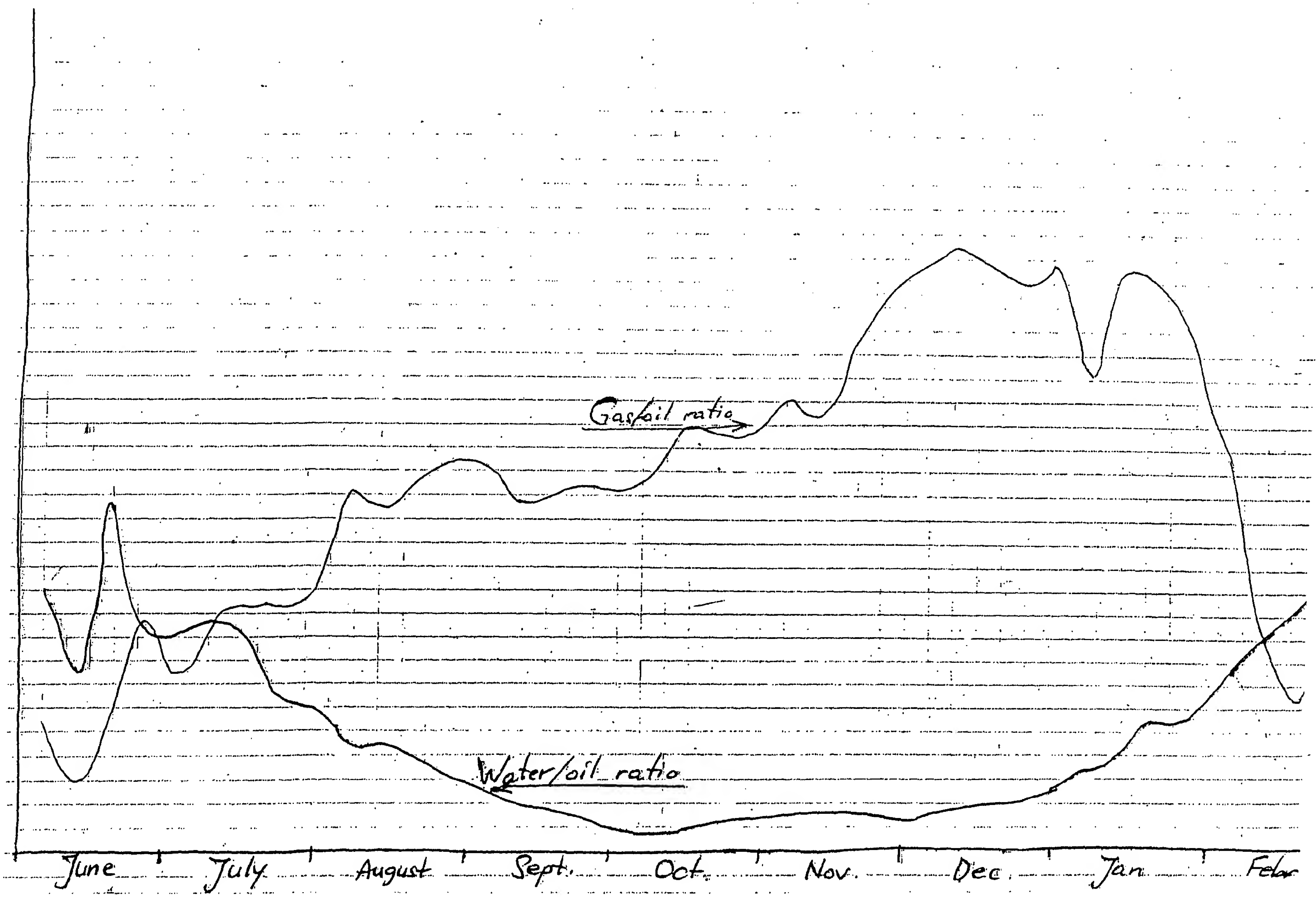
Summary

GOR	°API
760	25.8
1550	28.4
2470	30.0
3410	29.8
4380	29.2
5050	31.7

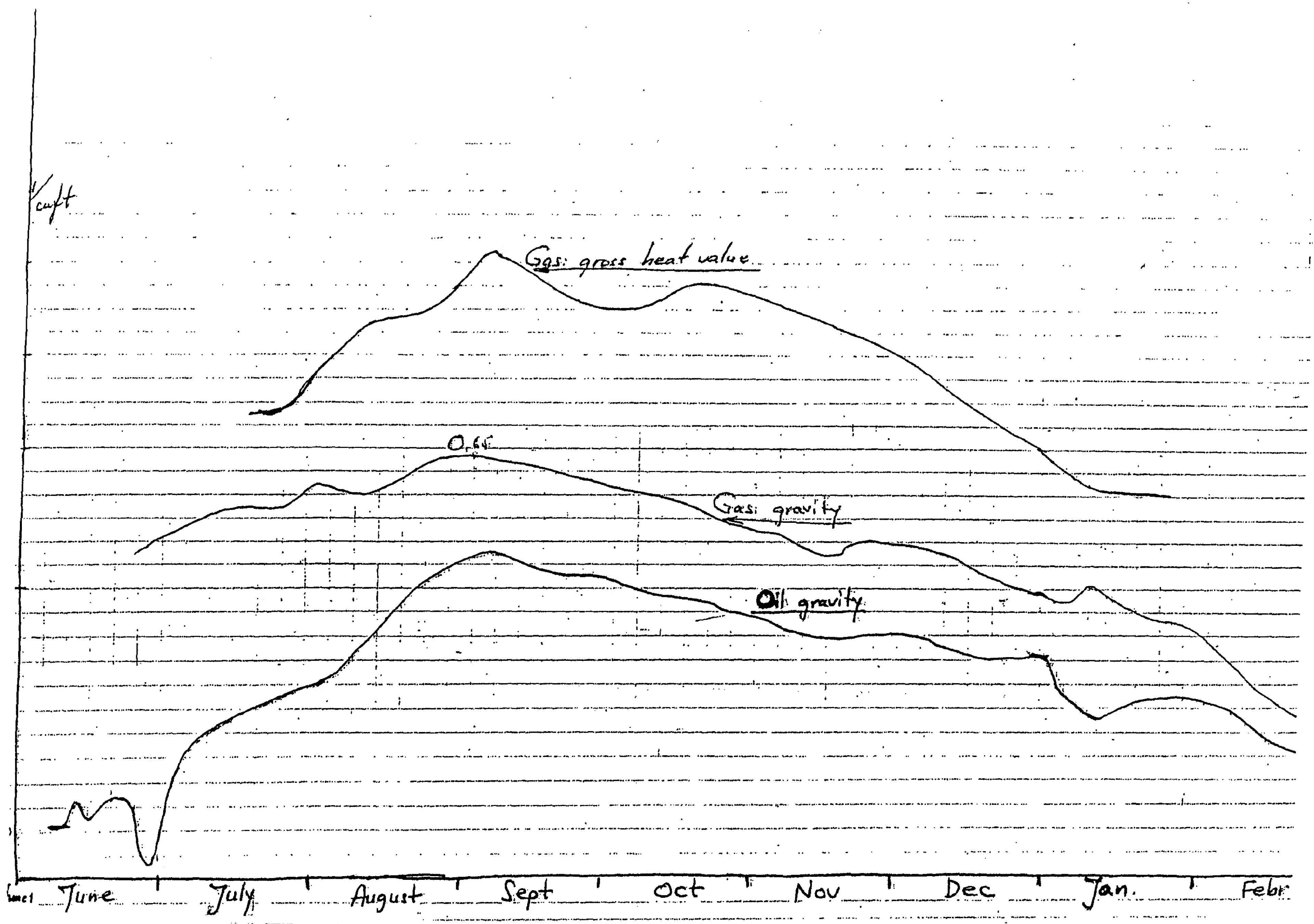




Water production
Gasproduction
Oil production
June
July
August
Sept.
Oct.
Nov
Dec.
Jan.
Feb



Gas/oil ratio
Water/oil ratio
June
July
August
Sept.
Oct.
Nov.
Dec.
Jan.
Feb.



/cuft
Gas: gross heat value
0.65
Gas: gravity
Oil gravity
June
July
August
Sept
Oct
Nov
Dec
Jan.
Febr.

Composite Analysis of L9 Produced Oil

Specific gravity at 60°F

°API	27.4
g/cm^3	0.8905
lbs/gallon	7.415

Gross heat of combustion at 60°F

10^6 BTU/bbl	5.96	
Characterization factor, K	11.5	
True Specific heat of oil at 65°F	0.457	BTU/lb, °F
" of oil 300°F	0.566	
Mean spec. heat 32°F – 300°F	0.494	
Mean spec. heat of oil vapor between 32°F and 300°F	0.404	

Latent heat of oil vapor at 300°F 94.3 BTU/lb.

Test nr.	Burner length ft	Heat input MBTU/hr	Sand filling: quant. ft of casing	Sand filling: size "mesh"	Sand filling: sand loss ft/day	Temp. readings: after 10^6 BTU supplied	Temp. readings: core T_{max} °F	(along casing) $\frac{T_{avg}}{T_{max}}$ %	L_{80} ft	L_{50} ft	Remarks
108A	5	20	0	—	0	1.4	1145	89	5.5	6.5	
108B	"	"	1.2	60-100	0	4.3	1050	98	7	10	
108C	"	"	1.2	"	0	2.4	1090	95	7	10	} 15 ft extension
108D	"	"	2	"	0	4.3	785	98	7	11.5	} tube ab. core
106A	10	20	0	—	0	0.6	1030	51	2	6	
106B	"	"	2	60-100	0	4.8	720	87	14	17.5	
113B	"	25	2.5-5	40-60	1.3	18	750	92	11	16	
113A	"	30	5	"	0.9	1.8	625	94	11	17	
107A	15	20	0	—	—	1.0	795	33	2	6	
107B	"	"	2	60-100	0	2.1	1190	37	2	4	
107D	"	"	1.5-3	"	0.66	1.0	1090	57	3	9.5	
115A	"	"	5	40-60	0.01	[illegible].5	420	81	13	15.5	
116A	"	"	5	"	0.02	24.3	405	90	16	20	} 20 ft ext. tube
115E	"	"	10	20-40	0	1.0	405	82	13	21	
115C	"	25	7	40-60	0.11	5.5	690	85	17	20	
115F	"	"	7-10	20-40	0.27	3.7	520	89	23	31.5	
107F	"	30	1.4-2	16	0.16	8.5	960	83	11.5	23	
107E	"	"	1.5-3	60-100	0.45	2.3	1010	74	6.5	22	
107C	"	"	2	"	0	0.8	790	77	6.5	20.5	
107G	"	"	2.2-3	40-60	0.11	5.1	905	90	16.5	23.5	
118B	20	20	9-10	10-12	0.25	6.7	445	87	24.5	33	
118A	"	22	8-10	20-40	0.13	7.9	460	94	33.5	36.5	
118C	"	25	8.2-10	10-12	0.27	6.1	575	92	32	37.5	
118D	"	30	6.5-10	10-12	0.78	13.1	620	91	33.5	38.5	
118F	"	"	7.5-10	8-12	0.57	9.5	625	96	31	35	
117A	"	"	9-10	8-12	0.54	15.9	860	71	29	34	
116E	25	20	9-10	10-12	0.07	6.4	575	78	19	34	
116A	"	25	5.5-10	40-60	0.43	11.6	450	86	28	33.5	
116F	"	"	8-10	12-14	0.19	21.1	590	93	32	39	
119A	"	30	9-10	12-14	0.58	8.6	660	86	28	40.5	
119B	"	"	"	8	0.38	14.3	725	84	26.5	40.5	
111A	"	21	10	40-60	0	20.2	315	78	15	26	15 ft ext. tube
111B	"	27	6-8	40-60	0.62	23.5	445	90	31.5	39.5	"
111C	"	"	7-9	10-30	0.14	2.1	500	79	24	30	"
111F	"	32	5.5-10	20-40	0.77	16.9	540	90	34	40	"
111D	"	"	7-8	10-30	0.14	3.2	600	86	24	34.5	"
111E	"	38	6-7	10-30	0.30	3.7	725	85	24	36	"
109E	"	30	8	40-60	0	2.4	580	45	5	10	3½" casing
109D	"	40	7-8	40-60	0.14	34.2	470	78	7	28.5	"
109B	"	"	10	"	0	2.2	235	64	9	27	"
109C	"	50	6-8	40-60	0.78	6.1	530	72	6	26	"
112B	35	25	8-10	40-60	0.24	10.0	410	73	16	31	

Summary of test results

1. Average of series (each series consisting of same four combinations of sand filling and (mass) flow rate.)

Test series #	Burner diam, in d	length, ft L	Casing diam D	Ratio = R $\frac{L}{\frac{\pi}{4}(D^2-d^2)}$ ft/in²	Burner efficiency: Temp. characteristic $\frac{T_{avg}}{T_{max}} \cdot 100 =$ [illegible]%	Relative heated length $\frac{L_{80}}{L} \cdot 100 =$ [illegible]%	with 4-65 1-38
R 120	1.660	36	3.068	6.92	86.8	120	121
R 121	1.660	46	3.068	8.83	75.8	66	68
✓ 121-B	1.315	46	3.068	7.60	72.0	75	74
R 122	1.660	53.3	3.548	6.92	66.0	77	74
✓ 126	1.900	53.3	4.026	5.40	85.3	120	121

The two burner efficiency coefficients are assumed to be functions of d, L, D and R.

probably Thus, the found data correspond to these equations:

$$\alpha = +149 \cdot D - \underline{\underline{\overset{?}{121} \cdot d}} - 5.17 \cdot L + 21.2 \cdot R - 132$$

and

should be -65 $$\beta = 122\,D - 29.5 \cdot d - 6.00 \cdot L + 3.18\,R - 38$$

Conclusions: Favorable is:

1) Wide casing (big D)
2) Thin burner (small d)
3) Short burner (small L)
4) Big R. ?

2. Average of series ↑

Test series #	Sand-filling % of annulus	Mass flow rate lb/ft², sec	Burner characteristics	
			$\frac{T_{av}}{T_{max}} \cdot 100 = \%$	$\frac{L_{80}}{L} \cdot 100 = \%$
N - 1	40	0.233	79.2	79.7
N - 2	40	0.291	76.6	105.6
N - 3	50	0.233	73.6	82.5
N - 4	50	0.291	79.2	97.1

Evidently neither $\frac{T_{av}}{T_{max}}$ nor $\frac{L_{80}}{L}$ is a linear function of sandfilling and mass flow rate.

Average Sand loss lb/day	%/day	Test #	Burner description: burner diam. d inch	burner length L ft	casing diam. D inch	length annulus ft/inch²	Sand filling % of annulus	Gas flow: M = lb/ft², sec	BTU/h	Temp. curve characteristics: $\frac{L_{80}}{L} \cdot 100\%$	$\frac{T_{av.}}{T_{max.}} \cdot 100\%$	Remarks: $\frac{\Delta t_H}{\Delta t_H} = \%$ α_H	$\frac{\Delta t_L}{\Delta t_H} = \%$ α_L
7.5	1.2	120-1	1¼"	36	3	6.92	40	0.231 ~~0.233~~	39,630 ~~40,000~~	110	86	83	83
1	0.8	-2	"	"	"	"	40	.293 ~~.291~~	50,340 ~~50,000~~	146	94	95	90
5	3.2	-3	"	"	"	"	50	.235 ~~.233~~	40,380 ~~40,000~~	103	84	108	86
	2.6	-4	"	"	"	"	50	.296 ~~.291~~	50,710 ~~60,000~~	121	83	106	86
0.6	0.4	121-1	1¼"	46	3	8.83	40	0.236 ~~0.233~~	40,600 ~~40,000~~	54	77	86	86
5.5	3.5	-2	"	"	"	"	40	.296 ~~.291~~	50,940 ~~50,000~~	78	79	105	83
4.0	2.1	-3	"	"	"	"	50	.239 ~~.233~~	41,180 ~~40,000~~	54	72	114	82
5.5	2.8	-4	"	"	"	"	50	.294 .291	50,620 ~~50,000~~	74	75	111	82
		121-1B	1	46	3	6.00	40	0.233	46,200	63	79	82	76
		-2B	"	"	"	"	40	.291	57,700	57	52		
		-3B	"	"	"	"	50	.233	46,200	82	74	116 Sweet	64 gas
		-4B	"	"	"	"	50	.291	59,000 ~~57,700~~	96	83	73 big sand	189 loss
		-5B	"	"	"	"	50	.303	60,000	=	—	—	—
		-6B	"	"	"	"	50	.253	50,000	61	74	106	78
lb/day	%												
4.2	2.1	122-1	1¼	53.3	3½	6.92	40	0.237 ~~0.233~~	57,440 ~~58,500~~	71	70	79	67
0.8	0.4	-2	"	"	"	"	40	.295 ~~.291~~	74,080 ~~74,000~~	88(84)	66	(73)107	68
8.4	3.4	-3	"	"	"	"	50	.236 ~~.233~~	59,230 ~~58,500~~	76	62	125	62
10	4.0	-4	"	"	"	"	50	.297 ~~.291~~	74,660 ~~74,000~~	92	66	121	65
Ave. Sand loss inch/lb/day	%												
20"	6.8	126-1	1½	53.3	4	5.50	40	0.239 ~~0.233~~	77,700 ~~75,800~~	118	84 (60)	88	83
12"	4.7	-2	"	"	"	"	40	.294 ~~.291~~	95,500 ~~94,700~~	159	92	86	95
6.7"	2.1	-3	"	"	"	"	50	.237 ~~.233~~	77,120 ~~75,800~~	101	76	106	81
7.4"	2.3	-4	"	"	"	"	50	.293 ~~.291~~	95,350 ~~94,700~~	103	89	82	85

		Test #	Burner description: diam. d in	length ft	casing diam. D in	length annulus ft / D in	Sand filling % of annulus	Gas flow: M = lb/ft², sec	BTU/h	Temp. curve characteristics: $\frac{L_{80}}{L} \cdot 100 = \%$	$\frac{T_{av}}{T_{max}} \cdot 100 = \%$	Remarks: $\frac{\Delta t_H}{\Delta t_H}\%$	$\frac{\Delta t_L}{\Delta t_H}\%$
~ l/day	%/day												
	2.3	122-1B	1	53.3	3½	6.24	51	0.238 ~~0.233~~	66,660 ~~65,000~~	49	52	100	31 too big sand Con
		-2B	"	"	"	"	51	.291	82,100	—	—		
14	4.4	-3B	"	"	"	"	64	.236 ~~.233~~	65,925 ~~65,000~~	75	55	116	39 too big sand Con
		-4B	"	"	"	"	64	.291	82,100	—	—		
8	3.1 ~~[illegible]~~	127-1	1½	30	4	5.50	59	0.237 ~~0.233~~	77,220 ~~75,800~~	156 180	98 95	97 [illegible]	101
5.6	3.2 %/day	-2	"	"	"	"	"	0.238	77,580	143	93	[illegible]	23
Tests with sweetened product gas:													
		121-1S	1¼	46	3	6.92	40	0.233	40,880 ~~40,000~~	20	67	83	83
		-2S	"	"	"	"	40	.291	41,000 ~~50,000~~	30	73	98	87
		-3S	"	"	"	"	50	.233	28,600 ~~40,000~~	33	71	85	80
		-4S	"	"	"	"	50	.291	37,000 ~~50,000~~	50	72	100	79

Tests with longer burners (68 and 87 ft) were cancelled because of lacking equipment. (Hoist could not be used to lift longer burners than 53 ft.)

#	burner diam. d, inch	burner length L, ft	casing diam. D, inch	length ÷ R annulus ft/in²	% of annulus	lb/ft², sec	BTU/h	$\frac{L_{go}}{L} \cdot 100\%$	$\frac{T_{av}}{T_{max}} \cdot 100\%$	
115A	1	15	2½	5.65	50	0.230	20,000		13	
110A	"	"	"	"	50	.230	20,000		16	duplicate w. 25ft extens. tube
115E	"	"	"	"	100	.230	20,000		14	
115C	"	"	"	"	50	.290	25,000		21	
115F	"	"	"	"	80	.290	25,000		22	
118	"	20	"	7.52	60-75	.345	30,000		30	
	"		"							
116A	"	25	"	9.40	36-48	.290	25,000		27	15 ft exten. tube
111B	"	"	"	"	36-48	.310	27,000		29	
116D	"	"	"	"	48-60	.345	30,000		3	
111D	"	"	"	"	42	.368	32,000		25	big sand loss (~10% per day)
111F	"	"	"	"	36-60	.368	32,000		21	
111E	"	"	"	"	36	.437	38,000		20	
112B	"	35	-	13.20	34-43	.290	25,000		15	5 ft exten. tube
B. Sandburner tests, Nov. 1957.										
115-H	1	15	2½	5.65	90-100	0.230	20,000		18	
-J	"	"	-	"	"	.290	25,000		19	
-K	"	"	"	"	"	.345	30,000		19	
118-B	"	20	"	7.52	67-75	.230	20,000		24	
-C	"	"	-	"	"	.290	25,000		31	
-D	"	"	"	"	"	.345	30,000		33	
116-E	"	25	"	9.40	54-60	.230	20,000		19	
-F	"	"	-	"	"	.290	25,000		32	
-G	"	"	-	"	"	.345	30,000		25	

Summary of earlier sojourner tests

A. September 1957. (1" burners)

Burner length ft	Heat Input BTU/h	Sand: Size mesh	Sand: Quantity ft	Sand: Loss %/Day	Length of zone with a temp in excess of: 70%	80%	90%	Remarks
15	20,000	40-60	5	0	14	13	8	
15	20,000	40-60	5	0	17	16	10	Duplicate with 25 ft extension tube
15	20,000	20-40	10	0	17	14	9	
15	25,000	20-40	5	3	23	21	9	
15	25,000	20-40	8	2	27	22	10	
20	30,000	20-40	8-10	2	34	30	16	
25	25,000	40-60	6-8	2	32	27	12	
25	27,000	40-60	6-8	4	33	29	23	Dupl. with 15 ft ext. t.
25	30,000	14-16	8-10	5	13	3	1	
25	32,000	10-30	7	2	31	25	14	
25	32,000	10-40	6-10	10	27	21	3	
25	38,000	10-30	6	10	29	20	5	
35	25,000	40-60	8-10	5	22	15	5	5 ft ext. tube

B. November 1957 (1" burners in 2½" casings, 9-10 ft sand, 12-14 mesh)

Burner length ft	Heat input BTU/h	Sand	Length of zone with a temp in excess of: 70%	80%	90%	
15	20,000	9-10 ft, 12-14 mesh	20	18	8	
15	25,000		27	19	14	
15	30,000		24	19	11	
20	20,000		30	24	17	
20	25,000		36	31	18	
20	30,000		35	33	17	
25	20,000		28	19	12	
25	25,000		37	32	21	
25	30,000		38	25	9	

1" burner tube in 30 2½" burner casing
9' - 10' sand, 12 - 14 mesh.
= 13,5 - 15' annulus

T	1" B-Tube Ft	INPUT BTU/h	Length in ft of the temp along the b-cas. above the following percentages of the temp. at the cone level deducted with 100° F.								
			70 %			80 %			90 %		
			Below Cone	Above cone	Total	Below cone	Above cone	Total	Below cone	Above cone	Total
15H	15	20,000	14	6	20	13	5	18	6	2	8
-		25,000	14	13	27	13	6	19	12	2	14
I		30,000	15	9	24	14	5	19	10	1	11
18B	20	20,000	18	12	30	14	10	24	10	7	17
C		25,000	20	16	36	17	14	31	13	5	18
D		30,000	20	15	35	20	13	33	11	6	17
16E	25	20,000	22,00	26	28	24	5	19	9	3	12
F	25,	25,000	25	11	37	23	9	32	19	2	21
G	30,	30,000	25	13	38	22	3	25	9	0	9

Nominal pipe dimensions, schedule 40 standard pipe:

nominal size, inches	outside diam, inches	inside diam, inches	wall thickness, in
1/4"	.540	.364	.088
3/8"	.675	.493	.091
1/2"	.840	.622	.109
3/4"	1.050	0.824	0.113
1"	1.315	1.049	0.133
1¼"	1.660	1.380	0.140
1½"	1.900	1.610	0.145
2"	2.375	2.067	0.154
2½"	2.875	2.469	0.203
3"	3.500	3.068	0.216
3½"	4.000	3.548	0.226
4	4.500	4.026	0.237
5	5.563	5.047	0.258
6	6.625	6.065	0.280

TESTS WITH SANDBURNERS.

1. Summary.
2. Purpose of the tests.
3. Testprogram.
    a. The layout of the tests.
    b. How the tests have been run.
    c. How the results have been evaluated.
4. Tests in 3" casing.
5. Tests in $3\frac{1}{2}$" casing.
6. Tests in 4" casing.
7. Discussion of the different tests.
8. Conclusions and suggestions.

REPORT ON SANDBURNERTESTS.

2. Purpose of the tests.

The purpose of the tests, described in this report was, as mentioned in the beginning of the report "Proposed sand-burner tests" by Robert E.Helander, May 16, 1958, to investigate the use of the sandburner in thicker and deeper formations, as well as to obtain data for optimum design of burner installations.

[illegible]

The proposed test program has been followed as far as it has been possibl . However, the ~~Table II in this~~ program, ~~in this report, Table~~ was revised, when it was found in the first tests, that the massflowrates of 0.269 and 0.342 lb/ft^2 sec. were too high. These rates were lowered to 0.233 and 0.291 lb/ft^2 sec. resp. Table /.

In addition to this program some more tests have been run in order to get more complete data, or to substitute for some tests which could not be run. Some tests were also run with sweetened produced gas instead of propane in order to determine if any variations could be observed in the temp. by using different fuel gases. The changes will be described under the different tests.

Three 12" holes were drilled about 20' apart, to 95', 135' and 170'. In these holes 10 3/4" casing were placed to the following depths: 90', 130' and 160'. In the 90' deep 10 3/4" casing a 90' iron casing, 3" in diameter was placed and on the side of this casing a 2" iron casing for temperature measuring. A 3½" burner casing was placed similarly in the 135' deep well and a 4" casing in the 170' deep well.

The empty annulus in the 10 3/4" casing was then filled with seasand in order to prevent convection.

<u>The burners.</u>

The burners were placed in the burner casings so the bottom of the burner tube was 3' above the bottom of the burner casing. It was placed like this in order to make it possible to observe temperatures below the bottom of the burner tube.

The burner tube was made of carbon steel pipe except for the 5 feet close to the cone which was made of 25/20 stainless steel.

The cones were the same kind as used in the test 19. They were made of cast stainless steel 25/12. This alloy is also called THERMALLOY "470" and was manufactured by ELECTRO ALLOYS DIVISION of American Brake Shoe Co.

The supply pipes from the top of the cone were made of 15' of 3/8" stainless steel pipe and from there up ½" iron pipe.

A schematic drawing of the set up is shown on Fig. 1.

The sand bed.

Hydrofracing sand of 10 - 12 Mesh was chosen for the tests. This sand is round shaped and consists mostly of pure quartz.

The fuel system.

The propane, which was kept at a constant pressure, regulated the pressure of the air by a demand type diaphragme regulator. The pressure of the propane and air could ~~by this~~ arrangement ~~be kept~~ the same.

The measuring equipment for gas and air.

The flow was measured in rotameters and regulated by needle valves placed after the rotameters. After the needle valves, the gases were mixed and supplied to the burner through a rubber hose.

The measuring equipment for temperature.

In the 2" iron casing for temperature measuring, as described above, a number of 12 gauge iron - constantan thermocouples were placed and fastened to a centrilized ½" pipe. The measuring points were placed 15 or 20 feet apart starting with the first point at the bottom of the temperature casing.

The ½" pipe with the attached thermocouples could be raised up to any level in between the spacing of the thermocouples, so the temperature could be measured at any depth along the burner casing.

The temperature was recorded with a 12 point Leeds and Northrup recorder.

any one of the running tests.

The tests have been run from 3 to 10 days each, or until two temp. curves with 24 hours intervall showed none or a small increase in the temperature.

8. Factors which have affected the heat input during the tests.

Temperature on the fuel gas.

The rotameters were calibrated for 70° F and for every 10° change from this temperature (70° F) the heat input will change 1 %.

9. Pressure of the fuel gas.

If the calibrations are made for 80 psig and higher pressure are used, the heat input will change 0.7 % for every psi difference.

The readings.

The rotameters could easily be read within ± 0.5 % of the full scale. In some tests, the floats in the rotameters have not been steady, because of the slugging in the sand bed, therefore the readings of the rotameters have not been accurate in these tests, but by taking high and low readings, satisfactory results were obtained.

Heat input.

A part of the temperature record from test 121-2 and 122-9 is shown on a photostatic copy (fig. /). The heat input, calculat d from the hourly readings of the rotameters, and corrected for temperature and pressure, is plotted on the temperature record.

Outside temperature.

The outside temperature has been varying between 10° - 95° F. The average daily variation has been between 50 - 70° F. For the

Sand level.

As a rule, the amount of sand has been measured every day and corrected. There is no correction for the sand that has stuck to the walls of the casings for longer or shorter time. These variations in sand level ~~have been up to~~ 2' of sand in casing. Most of the time this sand variation has been less than 1 foot of casing. No special tests has been made to see how a small change in sand level ~~will~~ affect the temperature~~s~~.

All the tests in the 120 and 121 series have been run according to the schedule. There have been no difficulties whatsoever.

The temperatures obtained from these tests are shown in Fig. L100-445 to L100-457.

Additional tests with sweetened production gas.

121-1S to 121-4S.

It was felt that a sweetened produced gas should be tried instead of propane in order to compare the two gases as fuel gases for the burners. For this reason the 121 series was run once more with produced gas ~~out of~~ the L9 test. It was sweetened with iron sponge and compressed before it was supplied to the burner. The heat value of this gas was, as to supplied to the burner, 1000 BTU/scf.

Additional tests with 1" burner.

Tests 121-1B to 121-4B.

The 121 serie was ~~run over again~~, but with a 1" burner instead of 1 1/4" burner. The reason for these tests was to see if the same heat distribution could be obtained by using a thinner burner tube.

Tests 121-5B and 121-6B.

Two more tests were run at 300000 BTU/h and 50,000 BTU/h. Of these two, the test 121-6B (50,000 BTU) had an excessive sandloss and was not completed.

Tests 120-4B to 120-6B.

In order to determine how ~~a~~ change of the heat input would influence the (temperatures recorded, a test was run with 4

52,700 and 50,810 BTU/h.each heat input was run over 24 hours. A 36' long 1" burner was used for these tests.

Test 120-93.

After these tests, one test was run with a baffle, placed on the $\frac{1}{2}$" supply pipe 25 feet from the top of the supply pipe. This test was run because it was felt that a more efficient curve was could be obtained if the slugging could be limited at a certain spot in the slugging zone.

Test 120-103.

The thermocouples were checked by this test. It was done so that which the burner was running the thermocouple were pulled and the temperatures were measured with a thermometer.

These tests were completed without any difficulty.

Tests 123-1 to 123-4.

123-1.

Could not be completed. ~~The amount~~ of sand was not enough to cover the cone at that heat input. ~~Very high~~ temperatures were recorded at the cone level. The thermocouples were damaged by this test and could not be repaired.

123-2.

The test was completed, but the temperatures had to be taken with thermometers.

123-3.

While running this test, the casing burned off and the test could not be completed. The casing burned off at the cone level because of the high temperature which was caused by an insufficient amount of sand.

122-1B to 122-4B.

The damaged part of the 3½" casing was replaced and the 68 feet 1¼" burner was replaced by a 33' long 1" burner. The change from 1 1/4" burner to 1" burner was made because of the results of a test to determine the highest input for a 1" burner. This test showed that a 1" burner could supply more than 35,000 BTU/h without blowing out the flame when the burner was raised up out of the sand. The amount of gas equal to 35,000 BTU/h was the upper limit for the measuring equipment.

The amount of sand in these tests was the same as in the test

temperatures were taken with a thermometer.

The tests 122-1B and 122-3B were completed.

The tests 122-2B and 4B could not be completed because of excessive sand losses.

Test 124-1 could not be run with 21'4" ~~xxx~~ of sand in the casing, therefore more sand had to be added. The temperature curve from December 15-16 was taken with 30' of sand in the casing.

For this reason the burner was shortened from 69' to 53' and these tests were called 126-1 to 126-4.

<u>Tests 126-1 to 126-4.</u>

When the burner was shortened, accidently no additional supply pipe was added, therefore the cone was placed at the same spot as in the tests 123, but the bottom of the burner tube was 15 feet higher than in the previous test. Because this fact at the time was unknown, the low temperature at the bottom of the casing could not be explained. After running the test 126-1 and 126-3 the burner was shortened from 53' to 30' and 23 foot of supply pipe was added. This test was called 122-1. ~~The same~~ low temperature at the bottom of the casing was noticed. It was then noticed that the burner was placed 19 feet from the bottom of the casing. The burner was lowered to 3' feet from the bottom of the casing and run with the same heat input as in 122-1, 77,000 BTU/h. A good heat distribution all the way down ~~could~~ now ~~be seen from~~ the temperature records.

The remaining two tests 126-2 and 126-4 were completed after the burner tube had been lengthened to 53 feet.

When these tests were completed, the thermocouples were pulled, and the temperatures taken with thermometer as a check on the

the test 126-2.

[illegible] of [illegible] [illegible] [illegible]

✓190: Thermocouples [illegible] 2 [illegible] [illegible] Water [illegible] in line and [illegible]. [illegible] [illegible] [illegible] [illegible] 100°F deducted from measured temps. 120-[illegible] and 122-2 done.

✓192: 120-2, 3, 122-1 done. Temp. [illegible] at 120-4 must be wrong. Will be checked with next [illegible].

✓193: 120-4, 122-23 done. No [illegible].

✓195: 121-1, 2, 122-4 done.

([illegible] next week)

✓197: 121-3, 4 done. Series 123 and 124 started. More [illegible] to be used than anticipated in program. Must be [illegible]. One in 123 burnt off. 121-S to be started next week.

✓199: Series 121-S started. Too high temperatures in 121-2S because of unstraight cone. Therefore shut off. (Top [illegible] [illegible] [illegible]) 123-2 and 124-1 running [illegible] [illegible] [illegible] 124 [illegible] will be cut from 68.5 to 53.5 ft [illegible] to overcome the difficulties. Temp. measurement troubles in 124-2.

✓200: 121-S tests restarted. 123-2 still running. 124-1 shut off.

✓201: 121-7S, 4S done. [illegible] [illegible] [illegible]. Series 123 starts to follow program. Series 125 and 126 [illegible] because of too short [illegible] to lift the burner.

✓202: 121-7S, 2S done. 123-2 done.

✓203: [illegible] 1, 123-1 [illegible], 126-1 [illegible] 124-1 [illegible] [illegible].

✓204: 123 [illegible], 126-3 going on.

✓205: 121-1S-4S done. 121-1B-4B being done ([illegible] instead of [illegible] [illegible] [illegible] with 15% higher (BTU/H)). Has been [illegible] that 30,000 could be supplied through 1" burner. Series 127 [illegible] substituted by 122-1B-4B. Test 127-1 identical with 126-1, but with only 20 [illegible] [illegible] of 53'.

✓206: 121-3B, 4B, 5B, 6B done, also 126-1, [illegible]. Tests 121-2B, 122-1B, 127-1 going on.

[illegible] equal to 20. [illegible], in [illegible]
of 126-1,3 because burner not sunk to bottom. Same with 127-1. Will be redone.

V208. 121-1B, 122-1B and 127-1(new) ~~soon~~ done.

V209 121-1B, 122-1B done. Slug preventer to be tested. Series 122B will take 5 more weeks. 127-1 being done, trouble with this.

V210: 127-1 (new) and 122-3B done.

V211: 122-3B, 126-4, 127-1 (new) done. 126-2, 120-4B being done. In latter ±10% input variation to be tested.

V212: 126-2 done.

A. ~~Fuel gas~~: propane, natural gas, produced gas, mixture
They have different flame velocity. If a burner operates well on propane, it will operate also on N-gas or P-gas. ∴ No problem

B. Size and type of sand: quartz sand 8-12 mesh to be used.
No reason to vary sand

C. Construction materials: ~~Tested~~ studied previously. No influence on burner performance.

Thus the following variables should be studied:

1. Heat input
2. Burner diam.
3. Burner length
4. Casing diameter
5. Thickness of interval
6. Height of sand

These were regrouped as follows:

1. Mass flow rate of gas (lb/ft² of annulus area)/sec)
2. ~~Sand height~~ / burner length
3. ~~Burner length~~ / annulus area
4. Casing diam.

} independent

5. Heat input
6. Thickness of heated interval

} dependent

1) First some trouble in avoiding condensing water in fuel lines and in rotameters. Also troubles with thermocouples.
When casing was too short, big sandlosses occurred.
When casing was too long, water condensed in upper parts and sand clogged! The sand clogging came loose occasionally and fell down in casing.

3) Cone in 123 burnt off once, and in 121-35 cone was not straightly welded to tubing.

4) Tests with burners, 68 ft or longer, cancelled because lifting hoist could not handle these lengths.

5) Slug preventers were tested without success.

Pressure drop in LINS burners

(Malte's test # 124)

1. 21'4" sand (casing)

Gas flow, stcuft		Pressure drop, psi: burner + hose	Pressure drop, psi: fluidized sand
140	126		4.5
214	190		8.0
267	253		10.8
320	316		13.4
350	384	1.5	14.5
446	448	3.5	14.5
575	575	4.5	16.5
580	583	6.0	17.5
646	650	7.0	18.8
687	723	9.0	19.1

When the F-line exploded Oct. 31, 1957, no remarkable conditions existed. The line had supplied sometimes more, sometimes fewer burners with air-propane-mixture, the pressures were about the same as normal, the temperatures have occasionally been as well higher as lower than on that day.

~~Among~~ The reasons which can be suggested now, are:

1) a blow on the line created a sudden pressure wave inside the pipe. (No work was going on close to the line)

2) a catalytically acting, igniting substance may have accumulated inside the F-system and caused the propane-air mixture to ignite, e.g. iron oxide or other substances, originating from the steel pipe, oxidation or other deterioration products from the rubber hoses, metals or other substances, originating from pipe fittings, valves, ~~gauges~~ air and propane tanks, gauges, rupture discs, flange jointings, deposits ~~from the~~ of lubricating oils from the compressors, oxidation and decomposition products of such lubricants, including coke, condensed water, impurities from the air, impurities from the propane (butane, pentane, sulfur compounds etc.), reaction products between air and propane (ketones, aldehydes, carbon monoxide, carbon dioxide etc.)

Gas explosioner och skydd däremot.
(Reserapport H571)

Den 21.11.57. besökte jag Bureau of Mines' Experiment Station, Pittsburgh, Pa., där jag sammanträffade med Dr. Dames (?) och Dr. M.G. Zabetakis, i Gas Explosions Branch, Explosives and Physical Sciences Division.

Jag redogjorde för de explosioner, som inträffat i våra luft-gas-ledningar i Santa Cruz. En skiss över situationen vid en typisk av de 6 explosioner, som inträffat, återges å bilaga 1. Ledningssystemet är normalt fyllt med en stökiometrisk blandning -

Union Oil Company of California

RESEARCH DEPARTMENT

BREA, CALIFORNIA

May 13, 1959 JES-66

Mr. M. F. Westfall (3)
Husky Oil Company
Cody, Wyoming

Dr. Gosta Salomonsson (3)
Svenska Skifferolje Aktiebolaget
Västra Gatan 2
Örebro, Sweden

Gentlemen:

Our laboratory analyst at the Shale Demonstration Plant has completed the Fischer assay tests on the 7 core samples submitted by Mr. Bengt Persson from the Swedish Process field test at Santa Cruz. The analytical data are contained in the attached table. No oil was recovered from any of the samples indicating they were quite well pyrolized in the formation.

If you need additional copies of the report, we shall be happy to supply them.

Very truly yours,

John E. Sherborne/RSC

John E. Sherborne, Manager
Production Research Division

RSC:vb
Attachment

cc/w: R. E. Helander
B. Persson
W. J. Shirley

UNION OIL COMPANY OF CALIFORNIA
SHALE DEMONSTRATION PLANT

OIL SHALE FISCHER ASSAY

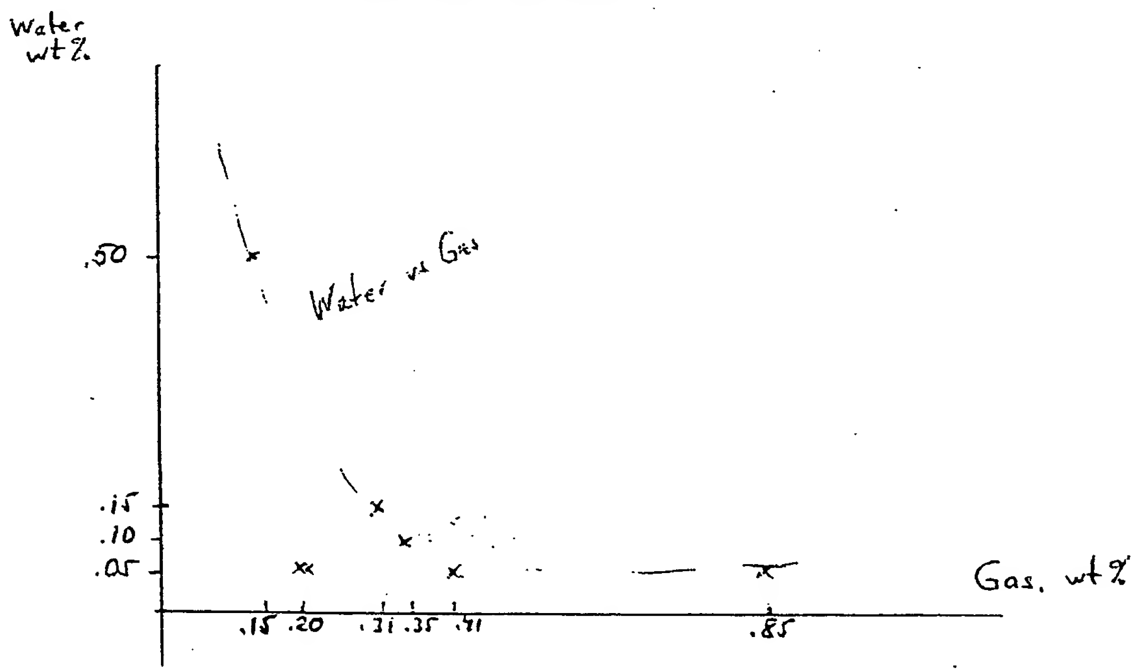
Date of Test: May 7, 1959

For: E. R. Atkins, Jr.

Assay Tests, Swedish Process Field Test
Core Samples from Santa Cruz, California

Sample Number	Sample Date	Yield, GPT Oil	Yield, GPT Water	Oil Wt. %	Water Wt. %	Residue Wt. %	Gas + Loss, Wt. %
C14-30-35	4-30-59	nil	0.4	0	0.15	99.54	0.31
C13-35-40	"	"	1.2	0	0.50	99.35	0.15
C11-30-35	"	"	0.1	0	0.05	99.75	0.20
C13-30-35	"	"	0.2	0	0.10	99.55	0.35
C9-20-25	"	"	0.1	0	0.05	99.10	0.85
C13-25-30	"	"	0.1	0	0.05	99.75	0.20
C14-15-20	"	"	0.1	0	0.05	99.54	0.41

Note: Samples analyzed in condition received



Enl. Huskeys juli-kostnader (1955):

28 ft, 1" i.d. 18-8-rör
70 ft 2⅞" i.d. 18-8-rör } kostade fr. Gate City Σ $ 1016:00
20 ft ¼" i.d. 18-8-rör

1008' 1" standard järnrör kostade $194,18 = 19,4 cts/ft
315' ¼" — " — " 26:– = 8,3 — " —
315 ¾" — " — " 45:– = 14,4 — " —

1 st. ½" globe valve kostade $ 2:48

1 st. ¾" — " — $ 3:28

Provkalkyl för en brännare genom 20 ft overburden + 80 ft tjärsand. (Delvis på amerik., delvis på svenska priser)

20 ft	¼" nedledningsrör	$ 1:66
1	jet + collar	0,50
4 ft	1½" hood, 18-8 à 4,00 $/ft	16:00
1	cone	1,00
6 ft	¾" 25-20 rör à 4,50 — " —	27,00
74 ft	¾" 18-8-rör à 2,50 — " —	185,00
1 st.	¼" needle valve	2,00
1	supply hose, 4 ft, with nipples	0,50
		$ 233:66

Do för gashål o. beklädnadsrör:

100 ft	3" järnrör à 0,60 $/ft	$ 60,00
25 ft	6" "	25,00
3 ft	3½" " (expansion tube)	2,00
4 ft	4"	0,80
1 st.	1" needle valve	4,00
		$ 92,00

1 brännare, anv. 4 ggr = ¼ · 233,66 = ~ ~$80:00
1 gashål + bekl.-rör, anv. 1 gg = ~ 100,00

Σ $ 180,00

Nelson sid. 189. För S-rika råoljor användes i raffinaderier rör och anordningar o dylika utrustning

0 - 290°C	0	5% Cr
230 - 430°C	5-8	7-12
370 - 480°C	7-12	12-18
480 - 540°C	5-8	12-18

Nelson sid. 190: Prisrelationer

0% Cr	—	11.8	$/enhet
5 —	—	3-3,5	
7 —	—	3,9-4,6	
9 —	—	4,2-5,3	
18	—	11-15	

Legeringar med >13% Cr förlorar sin seghet vid höga temperaturer, vilket motverkas med Ni-tillsats.

Nelson O.G.J. 1953:

H_2S-beständighet:	rel. H_2S-angrepp	rel. pris
kolstål	100	1.0
1% Cr	54	1.9
2	43	2.1
4	35	3.0
6	31	4.2
9	24	5.3
12	18	7.5
18	13	10.0
25	6	18.0

område					1200	1300	1400	1500	1600	1700
<1000°F	0	0	0	1.00	50	100	200	300	—	
<1000°F	0	0.5	0,1-0,5	1.53	0 "	"				
950°F – 1050°F	0,5	0.5	0.1-0.3	1.78	"	"				
ånga	1	0,5	0.5	1.86	"	"				
<1100°F ånga, olja	1,25	0.5	0.5-1,0	1.95	"	"				
<1150 olja	2	0.5	<0.5	2.10	"	"	"	"		
1175	2,25	1.0	<0,5	2.44	"	"	"	"		
1200	2,25	0,5	0,5-1.0	2.28	"	"	"	"		
1175	3	1,0	<0,5	2.49	—	—	—	—	—	
1200	5	0.5	<0.5	2.76	20	50	110	190	320	–
1400	5	0.5	1,0-2,0	2.99	0	0	0	0	55	115
1250	7	0.5	0,5-1,0	3.12	—	—	—	—	—	
1300	9	1	0,5-1,0	3.98	0	0	0	135	275	430
1300	12	(0,5Ni) 0	<0.75	7.70	0	0	0	95	205	310
50	12	(0,5Ni) 0,2Al 0	<0.75	7.70	0	0	0	0	110	230
1500	17	(0,5Ni) 0	<0.75	8.08	0	0	0	0	35	80
1600	18	(8Ni)	<0.75	8.76	0	0	0	0	(10)	(25)
2000	25	(12Ni)	<0.75	12.07	0	0	0	0	(10)	(25)
2000	25	(20Ni)	<0.75	17.09	0	0	0	0	(10)	(25)

Nelson, p. 198:

	% Cr	% Si	% Mo	% Ni	Arbetstemp. °F	Rel. oxidationsbeständighet (kolstål)
	0,5	1,3	0	0	1200	=
(Sicromal)	1	1,25	0,5	0	1200	1.2
(DM)	1,25	0.75	0.5	0	1200	1.3
(Croloy 2)	2	0,5	0,5	0	1200	1.5
	5	1,5	0,5	0	1300	=
(Croloy 9)	9	0	1,5	0	1300	17,0
	13			0	1400	84
(18-8)	18			8	1400	very high

SWEDISH EXTRACTION PROCESS

Santa Cruz, California.

Operate 2000 burners at 50,000 BTU:s/b-h, using 3½" OD burner casing, 1¼ burner tube, 18 ft sand, in an area having 60 ft 9 % tar sand with 50 ft over burden. Required fuel pressure 50 psig.

Tar 1 Acre = 43,560 x 60' x 115 x .09 = 27,050,760 #

Assuming an average gravity at the produced oil of 28.0° API

60% oil, and an oil recovery of 45 % by wt, oil recovery per acre would be 39,236 bbls.

60% = 52,400 bbls.

Assuming an average gas gravity of .666, and a gas recovery of 15 % by wt, gas recovery per acre would be 69,484 MCF (77,592). At an average heat value (after removal of H_2S) of 950 BTU/cu ft, heat value of produced gas per acre equal = 75,612.40 x 10^6 BTU

79,906

Assume 12' spacing, 124 · 68 ft^2/well or 349 wells per acre.

Heat required/Acre.

Heating time(hrs)	Lost heat	Million BTU/Acre Pyrolysis	Total	Input/ burner
3500	15050	54,886	69,936	57,254
3550	15156	54,886	70,042	56,533
4000	16090	54,886	70,976	50,842
4368	16790	54,886	71,676	46,559

Assume 6 month heating time.

vol. of air required = 2000 x 50,000 $\times \frac{1}{60} \times \frac{1}{100}$ = 16,667 cuft/min.

10 Fuller C 300 - 300 H blower required

Brake horse power/blower =

Compression ratio = 64.4/14.4 = 4.47 / Vol./day =

1700 x 60 x 24 = 2.448 mm cF

BHP/mm cF = 245 / .90 = 272

Area being heated = 2000/346 = 5.78 Acres.

$$\text{Daily gas vol} = \frac{\underline{69.484 \text{ MCF/Acre x } 5.78 \text{ Acres}}}{30 \text{ x } 6} = 2{,}231 \text{ MCF}$$

Gas vol/min. = 2,231,000/24 x 60 = 1549 cu ft/min.

H_2S contained in gas = 2,231 MCF x .127 = 283,337 cu ft =

283,337/60 x 24 = 196.76 cu ft/min.

Sulphur = 196.76 x .08515 = 16.75 #sulphur/min.

= 16.75 x 7000 = 117,250 grain/min.

Engine horsepower = 2721.8 1.96 = 354bhp

Air Compression.

I. Investment

1.	Fuller C-300-300H blowers	10	16940	$169,400
2.	Freight	190,000#	$4.84/cuft	9,196
3.	Waukesha VLROBU Engines	5	28117	140,585
4.	Freight	75,000#	$3.52/cuft	2,640
5.	After cooler	1		7,150
6.	Freight	13,400	4.84/cuft	649
7.	Pedestal bearings, belts and sheaves			7,500
8.	Concrete			3,000
9.	Setting			7,000
	Total			347,120

II. Monthly Operating Cost.

1.	Amortization 347,120/120	2,893
2.	Oil	360
3.	Maintenance	1,736
4.	Gas fuel	8,866
	Total monthly cost	13,855

Gas Sweetening and Compression Costs.

I. Investment.

1.	Fuller C-300-300H blower	1	$17240	$17240
2.	Freight	19000#	$4.84/cuft	920
3.	Waukesha LRORBU	1	13570	13570
4.	Freight	12000	3.52/cuft	422
5.	Aftercooler	1	1430	1430
6.	Freight			68
7.	Concrete			250
8.	Labor			500
9.	Erection and piping			25000
	Total			119124

II. Monthly Operation Cost.

1.	Amortization 119,124/120	993
2.	Oil	75
3.	Maintance	596
4.	Gas fluel	1074
	Total cost	2738

Labor Costs/mo.

1.	Project supervisor		750
2.	Engineer		500
3.	Head burner operator	40 x 4 1/3 x 2.70	468
4.	Burner Operator No. 1	40 x 4 1/3 x 2.54	440
5.	" " No. 2	374 hrs x 2.33	871
6.	Maintenance man No. 1	2 x 40 x 4 1/3 x 2.40	830
7.	Maintenance man No. 2	6 x 40 x 4 1/3 x 2.18	2265
8.	Welder No. 1	40 x 4 1/3 x 2.75	476
9.	Tester & Chemists Assist.	40 x 4 1/3 x 2.27	393
			6993

Assume an area covering 40 acres - 1320′x 1320′

Total vol. of fuel - 16,667 + 1,667 = 18,334 cu ft/min.

<u>Fuel system</u>.

1.	16" OD c .188" wall spiral weld csq	3000′		
2.	2 3/8"OD,3.75 L.W. T & C LP	20120′		
3.	2" couplings	80	59.57/100	172
4.	2" valves	80	15.12	1210
5.	½" couplings	4000	.1061	424
6.	½" Unions w/orifice plate	4000	.3144	1258
7.	½" ells	4000	.16	640
8.	½" x 2" nipples	8000	.07	560
9.	½" unions	4000	.15	600
10	½" rubber hoses	40000	.3533	14132
11.	Hose couplings	8000	.17	1360

12.	Gas air mixing equipment	2000
13.	Calorimeter	5000
	Total	

Product gathering, treating and storage.

1.	6 5/8" OD x .188" Armco csq	1500'		
2.	2 3/8" OD 3.75 L.W.T&C LP	20120'		
3.	2" couplings	80	2.15	172
4.	2" valves	80		1210
5.	½" couplings	8000	.1061	849
6.	½" Unions	4000	.15	600
7.	½" ells	4000	.16	640
8.	2" x ½" swage	4000		4760
9.	Heat exchangers			40576
10.	Pump			1000
11.	Treaters	2	7070	14140
12.	Tanks	4	2669.36	10677
13.	Stairway and walk way			487
14.	Misc piping and connections			2000
	Total			

Drill wells.

1.	Rig time	3.75 hrs	11.22	$42
2.	Bit cost	110'	.187	$21
	Total /well drilled			$63
1.	Rig time coring	8	11.22	$90
2.	Core head & core bbl repair 60 ft		.50	$30
				$120.-

Complete 367 holes/mo - core 10 %

core 37 holes	120/hole	4440
drill 330 holes	63/hole	20790
		25230

Oil wt/day = $\frac{27,050,760 \times .45 \times 5.78}{180}$ = 390,883
= 16,287 #/hr.

Gas wt/day = $\frac{27,050 \times .15 \times 5.78}{180}$ = 130,294
5,429 #/hr.

Water wt/day= $\frac{27,050,760 \times .10 \times 5.78}{180}$ = 86,863
3619 #/hr.

½ oil is vapor.

zone ~~2~~ - cooling vapor, condensate, gas and steam from
350°F to 210°F.

Cooling oil - condensate

8.143#/hr x 350 - 210) x .55 = .627 x 10^6 BTU

Cooling oil - vapors

8.143#/hr (350 - 210) x .45 .513 x 10^6

Condensing oil vapors

8.143#/hr x 120 .977 x 10^6

Cooling gas

5429#/hr x (350 - 210) x .57 .433 x 10^6

Cooling steam

3.619#/hr x (350 - 210) x .5 .253 x 10^6

zone 2 - Condensing steam at 210 F

3619 x 972 - .18 x 5429 x 1008 = .985 x 10^6

zone 3 - Cooling oil water and gas from 210°F to 150°F.

Cooling oil

16,287 x (210 - 150) x .5 .489 x 10^6

Cooling gas

5429 x (210 - 150) x .53 .173 x 10^6

Cooling water

3619 x (210-150) x 1.0 = .217 x 10^6

Total heat/hr 4.667 x 10^9 BTU

water required = $\frac{4,667,000}{8.33 \times 50}$ = 11,205 gal/hr.

Cooling surface = $\frac{2,803,000}{156 \times 49}$ = 366.6 ft^2

= $\frac{985,000}{117 \times 74}$ = 113.8 ft^2

= $\frac{879,000}{106 \times 30}$ = 276.4 ft^2

757 ft^2

Cooling oil from 150° to 80°F

16,287 x (150 - 80) x .46 = .524 x 10^6 BTU

Cooling surface = $\frac{524 \times 10^6}{28 \times 35}$ = 535 ft^2

water required = $\frac{524,000}{8.33 \times 20}$ = 3.145 gal/hr.

5429 (150 - 80) .5 = 190 x 10^6

5429 (0.18 - 0.03) 1025 = 835 x 10^6

1,025 x 10^6 BTU

$$\text{Cooling surface} = \frac{1{,}025{,}000}{28 \times 28} = 1.307 \text{ ft}^2$$

$$\text{water required} = \frac{1{,}025{,}000}{8.33 \times 20} = 6.152 \text{ gal/hr.}$$

Well equipment/well

1.	3½" OD 7.58 PE seamless LP	110′		
2.	3½" OD x .220 ASTM A-213-551 Tubing Gr 5 B	5′		
3.	2 3/8" OD 3.75 PE LP	50′		
4.	½" C.W. LP	67′		
5.	½" Couplings	7	.1061	.74

1.	11/4" x 25 - 20 stainless tubing	5′		
2.	11/4" x 18 - 8 " "	31′		
3.	Cast cone			7.50
4.	1/4" x H stainless pipe	10′	164.53	16.45
5.	1/4" x ½" stainless coupling	1	.45	.45

Water circulation

Thru compressor jackets

$$\text{air} = \frac{1700 \times 60 \times 10\ (360-100)\ .243}{8.33\ (120-70)} = \frac{64{,}443{,}600}{416.5} = 154{,}727 \text{ GPH}$$

$$\text{gas} = \frac{1700 \times 60\ (360-100)\ .55}{8.33\ (120-70)} = \frac{13{,}464{,}000}{416.5} = 32{,}327 \text{ GPH}$$

Thru After coolers.

$$\text{Air} = \frac{1700 \times 60 \times 10\ (100 - 80)\ .237}{8.33 \times (90-70)} = \frac{4,834,800}{166.6} = 29,020 \text{ GPH}$$

$$\text{Gas} = \frac{1700 \times 60\ (100 - 80)\ .53}{8.33\ (90-70)} = \frac{1,081,200}{166.6} = 6,490 \text{ GPH}$$

from product coolers = 20,502

Total water = 243,066 GPH
= 4,051 GPM

Investment

1. Centrifugal pump
2. hp electric motor
3. 16" OD x .188" wall Armco csq 1000′
4. Misc pipe and connections 1000
5. Cooling tower

Operating expense
1. Maintenance
2. Electric Power

Transportation

1.	Automobiles -	2 á 2500 mil/mo	8 c		400
2.	Pickup	2 " 250. "	10 c		500
3.	Trucks	1 " 176 hrs/mo.	5.00		880
					1780

Investment to start p 120 mo depr.

1.	Fuller C-300-300 H blowers (air)	10	16.940	169,400
2.	Freight	190,000#	4.84/cwt	9,196
3.	Waukesha VLROBU Engines	5	28.117	140,585
4.	Freight	75,000#	3.52/cwt	2,640
5.	After cooler	1	7.150	7,150
6.	Freight	13,400	4.84/cwt	649
7.	Pedestal bearings, shafts, belts, sheaves			7,500
8.	Fuller C-300-300 H blower (gas)	1	17,240	17,240
9.	Freight	19,000	4.84/cwt	920
10.	Waukesha LRORBU	1	13,570	13,570
11.	Freight	12,000#	3.52/cwt	422
12.	After cooler	1	1,430	1,430
13.	Freight			68
14.	Gas, sweetening, equipment			135,000
15.	Erection and piping			25,000
16.	16" OD x .188" wall spiral weld csq	4,000'	419.38/100	16,775
17.	Gas-air mixing equipment			2,000
18.	Calorimeter			5,000
19.	Heat exchangers			40,576
20.	Pump			1,000
21.	Treaters			14,140
22.	Tankage			11,164
23.	Misc piping and connections			10,000
24.	Centrifugal pumps	2	3,000	6,000
25,	75 Hp electric motor	2	2,000	4,000
26.	Cooling tower			40,000
27.	Labor			27,972
28.	Concrete			10,000
29.	$3\frac{1}{2}$" OD 7.70 CW T & C LP	12,000	88.74/100	10,649
30.	$4\frac{1}{2}$" O.D. 11.00 CW T&C LP	12,000	135.02/100	16,202
		Total		746,248

Cost/mo. = 746,248/120 = 6,219

Investment to start - 96 mo. Amortization.

1.	6 5/8" OD x .188" Spiral well scq	1,500′	210.40/100′	3,156
2.	2 3/8" OD 3.75 C.W. T&C LP	40,240′	43.78/100′	17,617
3.	Connections for fuel & Product lines			25,324
4.	½" LP	24,000′	9.88/100′	2,371
5.	1 1/4" x .140 25-20 stainless tubing	20,000′	421.55/100′	84,310
6.	1 1/4" x .140 18-8 stainless tubing	124,000′	315.11/100′	390,736
7.	Cast cones	2,000	7.50	15,000
8.	1/4" x H stainless pipe	40,000′	99.00/100′	39,600
9.	1/4" x ½" stainless coupling	4,000	.45	1,800
				579.914

Investment to start - 12 mo. amortization.

1.	3½" OD 7.58 PE Seamless LP	440,000′	92.96/100′	409,024
2.	3½" OD x .220 ASTM A-213-55TGR5B	20,000′	966.50/100′	193,300
3.	½" PE seamless LP	268,000′	11.28/100′	30.230
4.	2 3/8" OD 3.75 PE LP	200,000′	39.19/100′	78.380
				710.934

Total monthly costs

1.	Investment - 120 mo. amortization	6,219
2.	" 96 " "	6,041
3.	" 12 " "	59,245
4.	Labor	6,993
5.	Drilling	25,230
6.	Transportation	1,780
7.	Gas fuel	9,636
8.	Electric power	2,183
9.	Maintenance	5,000
10.	Oil	1.000
11.	Misc	5,000
12.	Sand	2,310
		130,637

Oil produced/mo 38,288 bbls.
Cost/bbl 3,41

Assume 13′ spacing, 146.38 ft^2/well or 298 wells/Acre

heat required/Ace

heating Time (hrs)	Lost heat	Million BTU/Acre Pyrolysis	Total	Input/burner
5000	17,988	54,886	72,874	48,909
5040	18,059	54,886	72,947	48,569

Assum 7 months heating time

Area being heated = 2000/298 = 6.71 acres

oil recovered/mo $= \frac{6.71 \times 39,236}{7}$ = 37,611 bbls

Total monthly costs

1.	Investment - 120 mo amortization	6,219
2.	" - 112 " "	5,178
3.	" 14 " "	50,781
4.	Labor	6,993
5.	Drilling	19,671
6.	Transportation	1,780
7.	Gas fuel	9,636
8.	Electric Power	2,183
9.	Maintenance	5,000
10.	Oil	1,000
11.	Misc	5,000
12.	Sand	2,310
		115,751

Cost/bbl = 115,751/37611 = 3.08

Assume 15′spacing, 195 ft^2/well or 224 wells/acre.

Assume 9.5 months heating.

Area being heated = $\frac{2000}{224}$ = 8.94 acres

Oil recovered/mo = $\frac{8.94 \cdot 39.236}{9.5}$ = 36,900 bbls

Total monthly costs

Investment - 19 mon.	36,500
Drilling	14,600
	51,100
Other items	46,162
	97,262

$$\text{Cost/bbl} = \frac{97,262}{36,900} = 2.64$$

Area per burner in triangular pattern = $\frac{3}{2}$ a^2

Area per burner in hexagonal " $\frac{3}{4}$ 3 a^2

$\frac{\text{Hexagonal p}}{\text{Triangular p}}$ = 1.5

Diagram 40 is for 1000 BTU/ft,h,b.
To heat up a deposit with 600 will thus correspond to a temp.
of $\frac{725 - 70}{600} \cdot 1000 = 1090^{\circ}F$

KL has 7.22 ft spacing which corresponds to 10.82 ft spacing with triangular pattern.

On the diagram multiply t e time with $\left(\frac{10.82}{7}\right)^2$ = 1.03 for the heaxagonal pattern in KL.

From the diagram a heating time of 3,550 hours =148 days =4.93 months is obtained for a hexagonal pattern of 7.22 ft or a triangular pattern of 10.82 ft.

The actual heating time in KL is 5.5 months. Thus the tarsand is heated up 11.5 % faster.

To heat 60 ft tarsand about 600 BTU/ft,hr should be delivered over 70 ft, 5 ft below and 5 ft above the tarsand. Through 45 ft of overburden about 100 BTU/ft-hr will be delivered. Thus 46,500 BTU/b-hr would be needed. However an additional 6.500 BTU/b-hr will be lost through the flue gas. Thus a total input of 53,000 BTU/b-hr will be required. 40 ft 1" burner tube in a 3" casing can be used.

A. Kostnaden för att tillföra berget en miljon BTU.

Värme tillföres berget genom förbränning av gas med luft i brännare, nedsatta i borrhål. Ett stort antal kombinationer av hålavstånd, brännareffekt och bränntid är tänkbara. I kalkylen nedan förutsättes att ungefär de förhållanden, som råder i Santa-Cruz-fältet tillämpas.

Priser och löner, gällande i California för närvarande, har använts. Räntan på investerat kapital antas vara 5 % per år och underhållskostnaderna för utrustning 4 % per år. Utrustningens livslängd är bedömd från fall till fall. Drifttiden per kalenderår antas bli 7900 timmar (90 % availability).

1. Borrhålet.

Borrning (inklusive rörsättning), 60 fot á 0,35 $	21.00 $/hål
Omborrning och uppdragning av ytterröret efter driftperiodens slut, 60 fot á 0,35 $	21.00 "
Cementering runt gasröret	2.00 "
Montagearbete (anslutning till ledningsnät för bränsle och pyrolysgas) (1 ~~timme~~)	2.00 "
Andel i kostnad för termometerhål (ett dylikt behövs för 20 - 100 brännarhål) ~65	1.00 "
Summa	48.00 $/hål

Antalet borrhål per acre beror på hålavståndet. Eftersom 57 · 10^9 BTU skall tillföras per acre (inklusive värmeförluster uppåt och nedåt), erhålles:

Hålavstånd, fot	8	10	12	15	20
hål per acre	790	500	349	223	126
borrhålskostnad, $/acre	38.000	24.000		10.700	6.000
" $/$10^6$ BTU	0,665	0,420		0,188	0,106

2. Rören.

jäm 80 cts/ft

Det har visat sig att rören kan upptagas och användas ånyo. 3 års genomsnittlig livslängd antages. Ytterröret antages vara av 5 % Cr, 0,5 % Mo, 1,5 % Si - kvalitet.

5Cr 0,5Mo, 1,5Si: 2.99$ ~~2.50 $~~/ft

20 fot gasrör (oleg.) á 0,80 $	16.00 $/hål	16
60 fot ytterrör (leg.) á 2.50 $	150.00 "	oleg. 48
Summa	166.00 $/hål	64 $/h

2,5 Cr 0,5 Mo, 0,75 Si: 2.28$ ~~1.85 $~~/ft

9Cr, 1 Mo 3.98$ ~~3.25 $~~/ft

Per drifttimme antages brännaren kunna inmata 25.000 BTU, varför kostnaden blir (med ränta, underhåll och avskrivning) 0,0083 $/drifttimme = 0.332 $/$10^6$ BTU.

-12
4.70 $/ft

Thermalloy 30 ~$13:-/ft

~~oleg. utan upptagning (10 fot hålavstånd)~~

$$\left.\begin{array}{l}\text{borrhål}\\ + \text{rör}\end{array}\right\} 701\ \$/\text{hål} = 0{,}89\ \$/10^6\ \text{BTU}$$

oleg. 1 års livslängd.

$$\frac{64(1.065)\cdot 10^6}{7900\cdot 25000} = 0.350\ \$/10^6$$

oleg. 2 års livslängd

$$\frac{64\cdot 0.565\cdot 10^6}{} = 0{,}182\ \$/10^6$$

3. Armatur, fasta ledningsnät m.m.

Andel i fasta förnät för tillförsel av

bränsle och bortförsel av pyrolysprodukter		15.00 \$/hål
kopplingar, ventiler etc.		5.00 "
	Summa	20.00 \$/hål

För dessa poster räknas med 10 års avskrivningstid, varför kostnaden blir 0,017 \$/10^6 BTU.

4. Brännaren.

Brännaren kostar, inklusive nedledningsrör och anslutningsdetaljer 52.00 \$/st.

Den antas kunna användas i 3 år med en inmatning av 25.000 BTU/drifttimme, varför kostnaden blir 0,096 \$/10^6 BTU.

5. Kompressorstationen.

En miljon BTU, tillfört tjärsandslagret, motsvarar ca 1,2 · 10^6 BTU i gasen eller 1330 cuft gas av värmevärdet 900 BTU/cuft (som gäller för såväl pyrolys- som naturgas). Motsvarande luftmängd är 12.000 cuft. Sammanlagt skall alltså 13.330 cuft gas + luft komprimeras till 12 psig (brännaren behöver 7 - 10 psig). Enligt kompressortillverkare kan man utan risk blanda gas och luft före kompressionen. En lämplig enhet skulle vara en kompressor med en kapacitet av ca 600 cuft/min, som räcker för 100 brännare à 25.000 BTU/h. En komplett enhet kostar:

kompressor		3.000 \$
elmotor (30 hkr) + varvtalsvariator		1.000 \$
blandningsregulator för gas - luft		700 \$
el- och gasledningar, fundament, montage		300 \$
	Summa	5.000 \$

Denna enhet antas ha 10 års avskrivningstid, varför den fasta kostnaden blir 0,105 \$/timme = 0,042 \$/10^6 BTU.

6. Kompressordriften.

Effektförbrukningen för en kompressorstation för 100 brännare är ca 18,5 kW, som vid kraftpriset 1,0 cts per kWh motsvarar 0,185 \$/drifttimme eller 0,074 \$/10^6 BTU.

Kompressorstationen kan göras praktiskt taget helautomatisk. Den tillsyn, som behövs, inkluderas i Arbetslöner.

7. Löner och administration.

Arbetsstyrkan för en 1000-brännaranläggning uppskattas bli 2 dagtidsarbetare (för underhåll) och 1 man per skift (för kompressor-, brännar- och pumpövervakning) För borrning erforderlig personal är inkluderad i borrkostnaden.

arbetare, 40 timmar/dygn à 2,00 $	= 80,00 $/dygn
arbetsledare (eller driftingenjör)	= 20,00 "
administration, 20 % av lönekostnaden	= 20,00 "
Summa	120,00 $/dygn

Kostnaden blir alltså 0,200 $/$10^6$ BTU.

Sammandrag vid hålavståndet	kostnad i $ per 10^6 tillförda BTU 8 fot	10 fot	15 fot	20 fot
1. Borrhålet	0,665	0,420	0,188	0,106
2. Rören	0,332	0,332	0,332	0,332
3. Armatur, ledningsnät	0,017	0,017	0,017	0,017
4. Brännaren	0,096	0,096	0,096	0,096
5. Kompressorstationen	0,042	0,042	0,042	0,042
6. Kompressordriften	0,074	0,074	0,074	0,074
7. Löner och administration	0,200	0,200	0,200	0,200
Summa	1,426	1,181	0,949	0,867

Anmärkning.

Det har här antagits att fältet är självförsörjande med bränslegas. Om så ej blir fallet kan tillsatsbränsle (naturgas) köpas för 0,50 $/$10^6$ BTU.

B. Oljeutvinningen per tillförd miljon BTU.

För att upphetta 1 cuft tjärsand till pyrolystemperatur åtgår teoretiskt 21,000 BTU. Om oljeutbytet är 4 vikts-% blir utvinningen 0,71 barrel per tillförda 10^6 BTU och om oljeutbytet är 6%, erhålles 1,08 barrel per 10^6 BTU.

I Santa Cruz-fyndigheten är genomsnittliga tjärhalten 8 vikts-%, varav man kan vänta sig att utvinna mellan 50 och 65 % som olja. För säkerhets skull räknes här med den lägre siffran, d.v.s. med 4 vikts-% oljeutbyte.

I ett enhålsförsök är värmeförlusterna till omgivningen mycket stora. Det kan matematiskt visas att endast 1,25 % av det tillförda värmet användes för verklig pyrolys. Sålunda erhålles per 10^6 BTU blott 0,0089 barrel. I enhålsförsök L 3 erhölls ca 0,02 barrels per 10^6 BTU, men tjärsanden var där rikare. (Den del av borrkärnan, som kunde tillvaratagas, höll ca 9% tjära.

I ett sjuhålsförsök är förlusterna till att börja med lika stora som i sju separata enhålsförsök, men efterhand som brännarnas samverkan kommer till synes, sjunker förlusterna, relativt sett, till ett minimum av ungefär 60 % av det tillförda värmet. Per 10^6 BTU erhålles då ca 0,28 barrels olja.

Efter lång tid flyter de sju brännarnas verkningar ihop till ungefär samma resultat, som skulle erhållas med en enda, sju gånger större brännare. Förlusterna motsvarar då ånyo förhållandena i ett enhålsförsök.

I försök L 72, där genomsnittliga tjärhalten var relativt låg, 7,3 %, erhölls totalt 4,16 barrels olja per $191 \cdot 10^6$ tillförda BTU eller 0,022 barrels/10^6 BTU. Korrektion till 8 % tjärhalt höjer siffran till 0,024 barrels/10^6 BTU.

I en mång-brännaranläggning beskriver de procentuella värmeförlusterna en liknande kurva som i en sjuhålsenhet med den skillnaden att minimiförlusten är konstant, så länge fältet kontinuerligt fortskrider framåt. Vid avslutning av ett begränsat fält stiger förlusterna åter.

För hundrahålsfältet L 8 har det beräknats att totalt 3400 barrels skulle erhållas med en inmatning av $11.900 \cdot 10^6$ BTU (fältets genomsnittliga tjärhalt = 7,3 %). Oljeutvinningen skulle sålunda bli 0,286 barrels/10^6 BTU. Under den tid fältet hade någotsånär konstanta driftförhållanden erhölls ca 0,09 barrels/10^6 BTU.

I en full-skala-anläggning med kontinuerlig fältflyttning beror förlusterna huvudsakligen på fältbredden och vandringshastigheten. I ett 2000 fot brett fält med 10 fots hålavstånd blir förlusterna ca 35 %, d.v.s. vid ett oljeutbyte av 4 vikts-% erhålles 0,46 barrels/10^6 BTU.

C. Sammanfattning.

De ovan gjorda kalkylerna visar sålunda att vid en fullstor anläggning med 10 fots hålavstånd tillverkningskostnaden för 0,46 barrels olja blir 1,18 $, eller för 1 barrel 2,55 $. Därtill skall läggas kostnaden för kondensering och lagring, som i en stor anläggning är blygsam, säg 5 cts/barrel.

Oljan skulle alltså kosta, fritt anläggningen 2,60 $/bbl.

För den olja, som hittills sålts, har erhållits 3,11 $/bbl. Den har emellertid varit något tyngre (spec.vikt 0,904) än vad som kan väntas från en fullstor anläggning (spec.vikt ca 0,880), varför försäljningspriset torde bli något högre. Transporten till kunden (raffinaderiet) kan väntas kosta max. ca 10 cts/barrel.

Kostnaden för gasens svavelrening har ej inkluderats i kalkylen, då den bör kunna bäras av det utvunna svavlet, för vilket ingen kreditering gjorts. Per m^3 olja blir svavelproduktionen av storleksordningen 30 kg.

Närkes Kvarntorp den 4 maj 1957

Gösta Salomonsson

Överingenjör

By heating tar sand "in situ" (without mining it) its content of tar is converted to gases, hydrocarbon vapours and carbon. The gases and vapours can be gathered through gas wells, while ~~the carbon~~ stays in the remaining sandstone.

The method has been tested at Santa Cruz. The two main problems which have been studied, are:

"How can the heat be supplied to the layer?" and "How much oil is recovered per supplied heat unit?"

The tests have not given the final answers yet, but the following conclusions may be drawn from the results hitherto obtained.

1. Heating.

The heat is supplied by burning a mixture of gas (~~—~~ produced gas or natural gas) and air in ~~a~~ burners in ~~a~~ drillholes, lined with a ~~[illegible]~~ steel casings, from which the heat spreads ~~[illegible]~~ to the surrounding rock. The burner is designed to ~~operate~~ require ~~at~~ not more than 10 psi air- and gas pressure and to supply from 15,000 to 30,000 BTU/hour.

The ~~[illegible]~~ spacing between the burners is determined mainly by economical considerations. The closer the spacing, the ~~[illegible]~~ smaller the amount of heat to be supplied by each burner, and the cheaper steel quality could be used for the casings (within some limits), but on the other hand the drilling cost and the number of

The optimum spacing at the Ljungström plant ~~i~~ in Sweden has been found to be about 7 ft.

The air and fuel gas are compressed to the required ~~[illegible]~~ pressure, either separately or after mixing ~~&~~ in a common compressor. The correct ratio air/gas is ~~[illegible]~~ automatically controlled. The most suitable unit size will probably be one mixing and compressing station for every 100 burners. If ~~the~~ the machinery is placed out along the field to be heated, ~~the~~ only power and fuel gas lines ~~[illegible]~~ to the stations are required (besides lines for fuel-air-mixture to each burner).

2. Heating costs.

This cost calculus is based on a heat unit of <u>10^6 BTU (= 1 MBTU), supplied to the tar sand layer</u>, whereby easy comparisons can be made between different field patterns. (The size of the unit is convenient also therein that <u>1 MBTU approximately produces 1 barrel of oil in a commercial field</u>. See below.)

a. Mixing station

1 MBTU, supplied to the rock, corresponds to about 1.2 MBTU as net heat value of the fuel or to 1330 stcuft of produced or natural gas, which both are assumed to have a net heat value of 900 BTU ~~[illegible]~~/stcuft. The combustion air amounts to 12,000 stcuft. Thus, a total of 13,330 stcuft of air + gas should be compressed to about 10 psig per 1 MBTU. The unit size for 100 burners,

= 30,330 stcuft/hour or about 560 stcuft/min. A mixing and compressing station for this capacity would cost

compressor .... 3000 $
electric motor (30 hp), incl. speed variator 1000 $
mixing equipment (controller) 700 $
connecting power and gas lines
and installation .... 300 $

Total 5000 $

If 10 years depreciation, 5%/year interest and 4%/year maintenance costs are assumed, the yearly costs will be:

$$5000 \cdot \frac{10 + 2.5 + 4}{100} = 825 \,\$,$$

or, with 330 days' availability per year:

$$\frac{825 \cdot 10^6}{330 \cdot 24 \cdot 100 \cdot 25{,}000} = 0.042 \,\$/\text{MBTU}$$

b. Compressor operation

The power consumption for compressing 560 stcuft/min from atmospheric to 10 psig pressure is about 25 hp or 18.5 kW. With a kWh-price of 1.5 cts, the power cost will be $18.5 \times 1.5 = 27.7$ cts/hour or $\frac{0.277}{2.5} = 0.111$ $/MBTU.

The attendance costs which are small on these highly automized units, are included in General Labor (see below).

c. Burner costs.

The cost of one LINS burner is $55.00 which includes the hose from the air-fuel supply line and 10

$3 \cdot 330.24 = 23.700$ hours of service. With 5 %/yr interest thus the burner cost will be $\frac{55 + 0.025 \cdot 3 \cdot 55}{23{,}700} = 0.00250$ \$/hour or $0.00250 \cdot \frac{10^6}{25{,}000} = 0.100$ \$/MBTU.

d. Well costs.

~~The~~ One burner well consists of a drillhole, 60ft ~~20ft~~ 4½" gas well casing, 60ft 2½" burner casing, a wellhead and a connection to the product line. In addition the well cost should include 5% of the cost for one temperature measurement hole (1 per 20 burner wells) and the cost for a well-distance length of the product line. The well-distance is assumed to be 10 ft. ~~Other distance~~ The Santa Cruz tests have shown that an alloy steel must be used for the burner casing, although it is still uncertain which quality should be used. It is here estimated that an alloy with 5% Cr 0.5% Mo 1.5% Si will be used. Further it is assumed that all casings can be pulled after use and used again for 3 years or 990 days of service.

I. Non-recoverable items:

60 ft drilling @ .35 \$/ft = 21.00 \$/well
(includes pipe-setting)

0.05 · 60 ft drilling for thermometer hole @ .35 \$/ft = 1.00 "

cementing gas well casing 2.00 "

washover drilling for pipe recovery
60 + 0.05 · 60 ft @ .35 \$/ft = 22.00 "
(includes pipe-pulling)

connection to product line, ...

Labor, 1 hour ... 2.00

Total 48.00 $/well

II. Recoverable items, 3 years ~~life~~ use

20 ft gas well casing, .80 $/ft 16.00 $/well

60 ft burner casing (5% Cr), 2.50 $/ft 150.00

~~Well head and connection and 10 ft of product line~~ ~~15~~ ~~20.00~~

0.05 · 60 ft thermometer casing @ .60 $/ft 1.80

Total 167.80 $/well

III. Recoverable items, 10 years use

Well head, connecting tubing and 10 ft of product line 15.00 $/well

5% thermometer stand with thermometer, chain and counter 2.00 (~~25.00~~)

Total 17.00 (~~40.00~~) $/well

With 5%/yr interest and 4%/yr maintenance costs on the recoverable items, the yearly costs for these will be:

$$\underbrace{\frac{167.80}{3} + 167.80 \cdot 0.065}_{66.70} + \underbrace{\frac{17.00}{10} + 17.00 \cdot 0.065}_{2.80} = 69.50 \ (\cancel{73.45}) \ \$/\text{year, well}$$

or

$$\frac{69.50 \ (\cancel{73.45}) \cdot 10^6}{\underbrace{330 \cdot 24}_{8000 \text{ hours}} \cdot 25{,}000} = 0.3495 \ (\cancel{0.367}) \ \$/\text{MBTU}$$

The cost of nonrecoverable items per MBTU can be calculated only after a certain well distance is chosen.

MBTU should be supplied (including vertical costs). Thus, if the number of burner wells per acre vary, the costs per MBTU will be as follows:

well spacing, ft	8	10	15	20
wells per acre	790	500	223	126
well costs, \$/MBTU:				
nonrecovable	0.665	.420	.187	.106
recoverable	0.367	.367	.367	.367
total	1.032	0.787	0.554	0.473

e. Labour and supervision costs

The total labour requirement for operation of a 1000 burner plant is estimated to 2 daytime workers (for maintenance) + 1 man/shift (for compressor, burner and pump attendance). In addition, one engineer is needed for supervision. Thus the labour and supervision costs will be: $2.00(24+16) + 20 = 100$ \$/day or

$$\frac{100 \cdot 10^6}{24 \cdot 1000 \cdot 25{,}000} = 0.167 \text{ \$/MBTU}$$

20% is added for payroll burden + overhead resulting in:

$$0.167 \cdot 1.20 = 0.200 \text{ \$/MBTU.}$$

(f. Fuel costs.

It is assumed that the field is selfsupporting with fuel gas. If this is not the case, the additional fuel can be bought as natural gas for .50 \$/MBTU)

	costs $ per MBTU			
well spacing, ft	8	10	15	20
wells per acre				
a. mixing and compression (10 years)	.042	.042	.042	.042
b. compressor operation	.111	.111	.111	.111
c. burners (3 years)	.100	.100	.100	.100
d. wells: recover. (3, 10 yrs)	.367	.367	.367	.367
nonrecover.	.665	.420	.187	.106
e. labour, supervision, overh.	.200	.200	.200	.200
Total	1.485	1.240	1.007	0.926

It is evident that the recovery and reuse of burners and well casings is of ~~the~~ utmost importance. If, for instance, the burners can be used for 3 years, but the well casings can be used in an average only 1 year, the costs will increase by 0.523 $/MBTU above the tabulated sums.

The losses to the surroundings around the field vary according to the volume of the field, compared to its heat-transferring border surface.

1. Theoretically the tar sand should be heated to 750°F, at which temperature the pyrolysis is complete. This requires a heat quantity of 21,000 BTU per cubicfoot of tar sand. Thus, if the oil yield is 4% by weight (corresponding to .015 barrel of oil per cubicfoot) the theoretical heat consumption is 1.40×10^6 BTU per barrel.

If 6% by weight is recovered the heat needed is 0.93×10^6 BTU and if 8% by weight is recovered, the heat consumption is 0.70×10^6 BTU.

2. In a single-burner unit the heat losses are tremendous and it can be shown analytically that only 1.25% of the supplied heat is used for actual pyrolysis. Thus 1 barrel of oil requires

112×10^6 BTU, if 4% by weight is obtained,
$74 \times$ " " 6% " " "
$56 \times$ " " 8% " " "

The single-burner test L3 produced about 2 barrels of oil after about 100×10^6 BTU had been supplied. Thus the heat consumption per barrel was about 50×10^6 BTU. The tar content of this area is not known exactly, because of partially lost core in the core drilling. The analyzed parts of the core indicate a tar content of about 9% by weight.

3. In a seven-burner unit the heat losses in the beginning are of the same order of magnitude as in a single-hole unit

isolated units. The heat consumption is then about 112 × 10⁶ BTU per barrel (at 4% b.w. recovery).

After a short time the interaction between the burners starts and a higher degree of efficiency is obtained. The heat requirement gradually decreases to about 3.5 × 10⁶ BTU per barrel (at 4% b.w. recovery), corresponding to about 40% efficiency.

After a longer heating period the actual zone, where the pyrolysis takes place has moved so far outwards that there is very little difference between the seven-burner unit and a single-burner unit with a sevenfold heat input per foot burner length. Thus the efficiency of the heating gradually approaches the 1.25%-limit again as an asymptote, or the 112 × 10⁶ BTU per barrel (4% b.w.) heat consumption.

In the test L72 4.16 barrels of oil were obtained after a heat input of 191 × 10⁶ BTU, corresponding to 46 × 10⁶ BTU per barrel. The average tar content of this area is about 7.3% by weight and if a recovery of 50% of the tar is assumed the figure 46 × 10⁶ BTU corresponds to 42 × 10⁶ BTU per barrel at 4% b.w. recovery.

In a multi-burner unit the efficiency of the heating developes the same type of curve as in a seven-burner unit, starting from 1.25%, rising to a maximum value, and then decreasing to 1.25% finally. The maximum figure, which remains valid during the main period of the operating time, depends upon the dimensions of the unit. For instance, in the 100-burner field, where the tar sand thickness is 45 feet and

0,000 cuft = 700,000 lbs

$10,280 \times 10^6$ BTU would be needed for the complete pyrolysis of the $45 \times 69 \times 80 = 250,000$ ~~cuft~~ cuft of tar sand, including losses to the surroundings. However, this figure was arrived at from the assumption a specific heat of .26 BTU/lbs, °F. It has been found later that this figure rather should be .30 (because of higher water content than anticipated) and thus the total heat consumption would be $\frac{.30}{.26} \cdot 10,280 \times 10^6 =$ $11,9000 \times 10^6$ BTU.

; 700,000 · 0.0365 = 1,050,000 lbs

$\frac{\text{~~265415~~}}{100 \cdot 7.3 \cdot 42} =$ 305 lbs/bbl

The oil production will be 3400 barrels, if 50% of the 7.30% by weight of tar in the sand is recovered. Thus the expected ~~efficiency of heating would be 3200~~ specific heat consumption would be $\frac{119,000 \times 10^6}{32,000} = \underline{\underline{3.7 \times 10^6 \text{ BTU/barrel.}}}$

$\frac{0,000}{305} =$

The oil production up to March 4, 1957 was 169.2 barrels and the heat input was 5177×10^6 BTU, corresponding to a specific heat consumption of $\underline{\underline{31 \times 10^6 \text{ BTU/barrel.}}}$

The result shows that the average heat efficiency is somewhere between the start figure of 112×10^6 BTU and the overall total of 3.7×10^6 BTU.

The present rate of production (during the last 3-4 weeks) has been 1.2 barrels/day and the heat input $32 \times 17,000 \times 24 =$ $= 13 \times 10^6$ BTU/day or $\frac{13 \cdot 10^6}{1.2} = \underline{\underline{\sim 11 \times 10^6 \text{ BTU/barrel.}}}$

<u>In a full-scale plant</u> on the same location as the 100-burner test (L8), assuming a continuous operation of a 200-burner wide field the steady-state heat consumption will be 2.6×10^6 BTU/barrel. Under the same conditions but with a recovery of 4.0% by weight instead of 3.65% by weight

Summary

unit size	specific heat consumption, 10^6 BTU/barrel	
	calculated	actually obtained
1-burner	112 (at 4% by wt. rec.)	~50
7-burner	start 112 — " — main ~~112~~ ~5 — " — final 112 — " —	42
100-burner	start 112 — " — main 3.7 — " —	accumul. 31 present rate 11
Continuous operation of 200-burner wide field	2.38 — " —	
Theoretical value (no losses)	1.4 (at 4% by weight recovery) .93 (— 6% — " —) .70 (— 8% — " —)	

1 gjuten kona, Thermalloy,	2.15	$/brännare
5 ft. 1"-rör, 25-20, à 4.35	21.75	— " —
10 ft — " — 18-8, à 1.30	13.00	— " —
40 ft ½" nedledningsrör, järn, 0.12	4.80	— " —
svetsning	5.00	— " —
styrfenor, koppling, styrbricka, rörböj, slang, nippel	5.00	— " —
Summa	51.70	$/brännare

2. Hål för brännare

borrning, 60 fot à ~~2 ..~~ 2x) (arbetslön)	9.00	$/hål
— " — à 0.35 (borrmaskin)	21.00	— " —
beklädnadsrör, 15 fot 4½" à 0.80 + bussning + cement + arbete	15.00	— " —
ytterrör, 60 fot, 2½" (5% Cr) à 2.65	159.00	— " —
"well lead" (bussning)	1.50	— " —
Al-rör, 4 fot ½" med anslutningar	1.50	— " —
svetsning av ytterrör	5.00	— " —
Summa	212.00	$/hål

3. Temperaturhål.

borrning, 60 fot	21.00	$/hål
— " — arbetslön	2.00	— " —
rör, 63 fot, 2" à 0.60	38.00	
Summa	68.00	$/hål.

Flyttbar T-håls-utrustning:

Skrivare	~~20~~ 25.00	$
Stållina	~~25~~ 50.00	"
Kabel 60 fot	9.00	"
Termoelement 3.85 $		
— " — hållare etc.	15.00	"
Summa	99.00	$/hål.

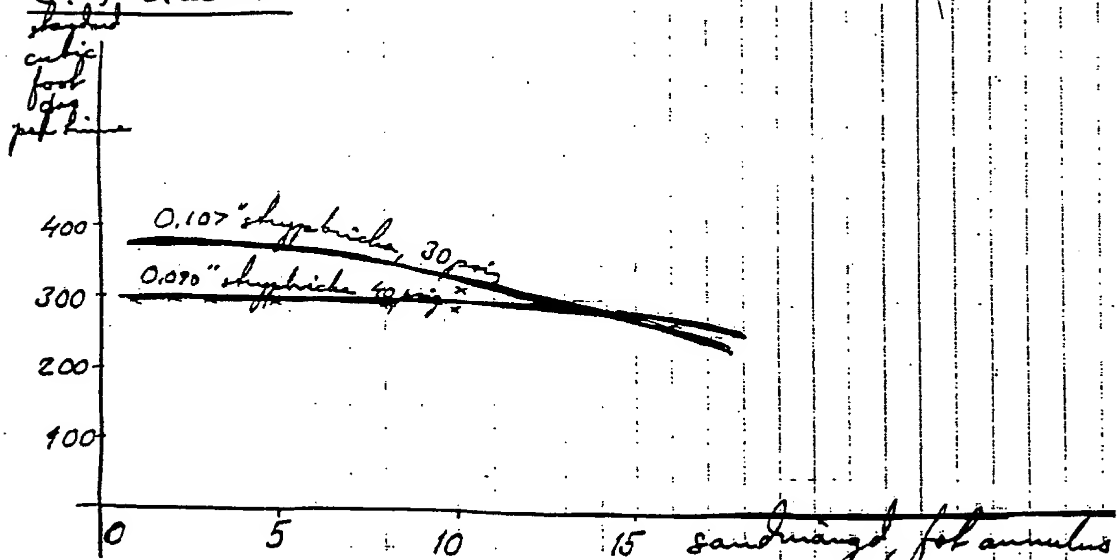
i en LINS-brännare.

(Sammandrag av M.E:s försök 118A, 3 juli 1958.)

1. Eftersom en viss sandförlust förekommer under drift av LINS-brännaren, är sandfyllningen ej konstant. Följaktligen varierar tryckfallet och därmed brännareffekten. Hur mycket?

2. Försök gjordes med en 20 fot lång 1-tums-brännare i ett 50 fot långt 2½-tums ytterrör. Varierande mängder 8-12 mesh Monterey-sand hälldes i ytterröret. I tillförselrörets lopp anbringades en 0,090" ~~strypbricka~~ resp. en 0,107" strypbricka. Bränsle-luftblandningens tryck före strypbrickan hölls vid 40 resp. 30 psig. ~~Mängd~~ Flödet mättes med kalibrerad rotameter.

3. Resultat



∴ Ju större andel av tryckfallet som ligger över strypbrickan, desto bättre konstanthållning av effekten.

Union Oil Company of California
RESEARCH DEPARTMENT
BREA, CALIFORNIA

April 22, 1959 JES-60

Mr. M. F. Westfall (2)
Husky Oil Company
Cody, Wyoming

Dr. Gosta Salomonsson (2)
Svenska Skifferolje Aktiebolaget
Västra Gatan 2
Örebro, Sweden

Gentlemen:

At one of the recent meetings of the Engineering Committee for the Swedish Process Field Test at Santa Cruz, we obtained a sample of burner casing recovered from one of the burner wells in the L-73 test. The casing had parted during the salvage operations and at the parting point the wall thickness had been reduced to a very small value. To determine the nature of the attack which occurred at this point we have had Dr. L. M. Dvoracek of our Design Division examine the specimen metallurgically. For your information, we have attached hereto copies of the report prepared by Dr. Dvoracek to cover his examination.

We believe the report is self explanatory; however, if you have any question regarding it, please contact us.

Very truly yours,

John E. Sherborne, Manager
Production Research Division

JES:vb
enc.

cc/w: B. Persson
R. E. Helander
W. J. Shirley

Union Oil Company of California

RESEARCH DEPARTMENT

BREA, CALIFORNIA

To:	Dr. Clyde Berg, Mgr. Design Division	Reference:	JEH-902M
		Date:	March 25, 1959
From:	Louis M. Dvoracek	Project:	62-11552
Subject:	TAR SANDS PROJECT	Supervisor:	John E. Hines, Jr.
cc:	E. R. Atkins (4) J. E. Hines J. R. Hunt		

HISTORY

Oil is extracted from tar sands by insitu heating. The heating is supplied by a pattern or network of combustion wells. These wells contain a burner inside a pipe or tube. Heat is transferred from the burner to a fluidized sand and thence to the wall or pipe of the well. The tar sands surrounding the wells are heated thereby releasing their oil. The carbon steel pipe housing this well is a 2-1/2-inch pipe of approximately 1/4-inch wall thickness. Normal operations of these wells are from 700°F to 1000°F.

agreed.

Removal of a well designed as L73, B1, failed at the 26-foot level. This well extends approximately 40 feet and has been in service for over a year.

EXAMINATION

Visual inspection at the failure area indicated very little parent metal in the order of 1/10-inch or less. Voluminous scale deposits were noted on both the inside and outside of the pipe. The outside deposits were greater than the interior and were black in appearance. An acid test of this outside layer indicates the formation to be largely sulfides, while the small reddish appearance of the inside layer to be oxides of iron.

Figures 1 and 2 reveal the character of the scale and mode of penetration. The parent metal is at the top of the photomicrographs. The demarcation between scale and metal is very uniform. Also indicated is a general structure simulating the parent metal. This might be the direct substitution or combining of sulfur and/or oxygen with iron atoms.

Figure 3 is a photomicrograph of the metal located about a foot above the failure area. The structure is largely ferrite with a small amount of pearlite. The carbon content is low, probably in the neighborhood of five points (0.05 percent).

A sharp contrast in microstructure is noted in the failure area as indicated by Figures 4 and 5. Figure 4 depicts the microstructure and boundary between metal and outside scale, while Figure 5 presents the interior portion of the tube. As already indicated, the penetration is uniform on both surfaces.

However, the structure from interior to the exterior position of the metal is very striking. Grain growth has occurred at the inside with the maximum size at the centerline of the wall. From the centerline to the exterior side, the grains are smaller but still possess growth. Pearlite is absent on the inside, but appears near the outside. The pearlite under higher magnification in this area is laminar. Carburization or decarburization is not visible. Apparently, the time-temperature relationships were high enough to produce grain growth with a high enough temperature to dissolve the pearlite on the inside portion of the wall which now appears in the spheroidized state. This means the temperature was in the critical range (1300°F). It is hard to visualize a temperature gradient across the wall, but apparently the time-temperature relationships were adequate for this effect. The hardness in this area was R_B 40. This is slightly lower than that reported in the literature for this type of carbon steel. Exposure to these temperatures will soften the material.

RECOMMENDATION

The attack to a carbon steel combustion well was severe on both the inside and outside of the pipe. Alloying with chromium will offer resistance to oxidation, sulfurization, and carburization. If the composition of the oil from the tar sands has considerable amounts of hydrogen and hydrogen sulfide, then alloy compositions of stainless steels would be required. However, high alloying is a costly solution.

Aluminum coatings or alloys also offer protection to oxidation and hydrogen sulfide attack. Coatings such as Metallizing, Mollerizing, or Calorizing offer great promise. Even the non-diffused aluminum coating which would become diffused or alloyed in service might prove to be very economical.

Louis M Dvoracek

Louis M. Dvoracek
Design Division

LMD:ef
attachments

Figure 1

Scale formed on the inside of combustion well L73, Bl. Etchant, Nital; 50X

Figure 2

Scale formed on the outside of combustion well L73, Bl. Etchant, Nital; 50X

JEH-902M
3/25/59
Research

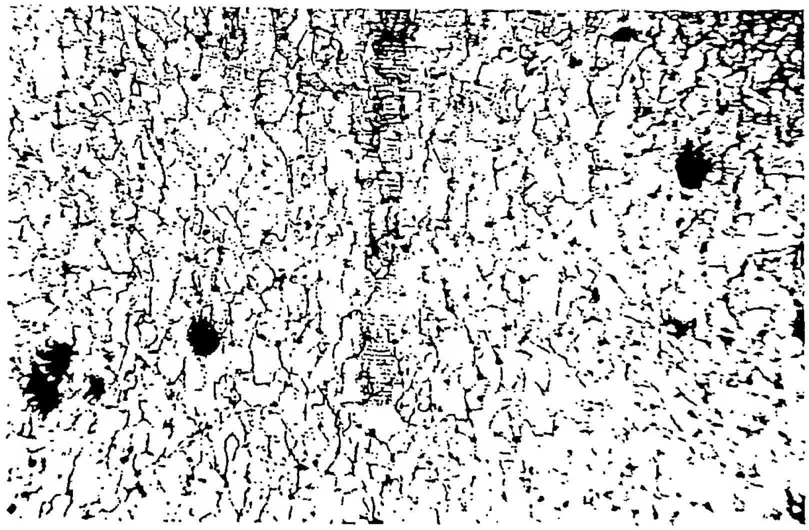


Figure 3

Microstructure of combustion well
pipe L73, Bl. Above the failure area.
Etchant, Nital; 150X

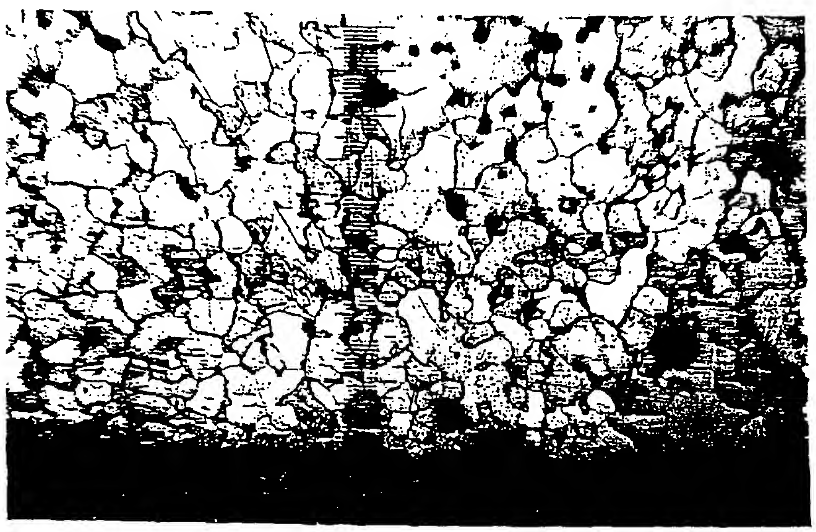


Figure 4

Microstructure along outside portion of
combustion well pipe L73, Bl.
Etchant, Nital; 150X

TEH-902M
3/25/59
Research

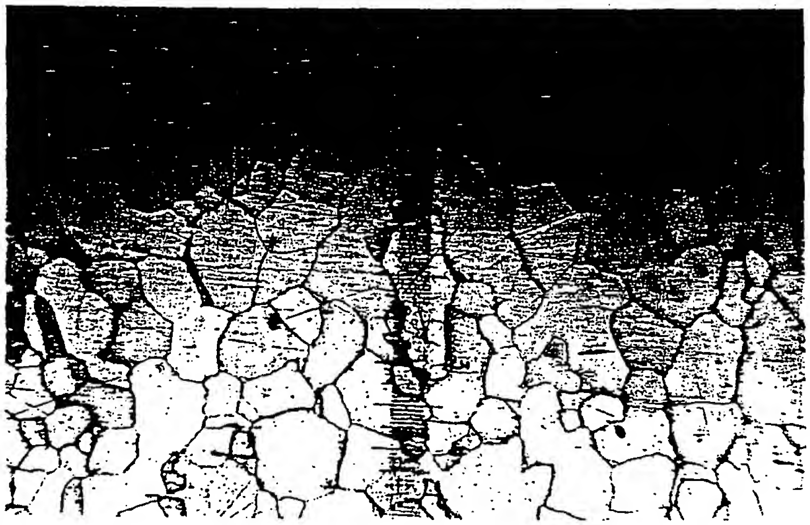


Figure 5

Microstructure along interior of
combustion well pipe L73, B1.
Etchant, Nital; 150X.

JEH-902M
3/25/59
Research

Union Oil Company of California

RESEARCH DEPARTMENT

BREA, CALIFORNIA

February 6, 1959 JES-16

Mr. M. F. Westfall
Husky Oil Company
Cody, Wyoming

Dr. Gosta Salomonsson
Svenska Skifferolje Aktiebolaget
Västra Gatan 2
Örebro, Sweden

Gentlemen:

We have completed various analytical tests on the L-73 core samples which we obtained during our last visit at Santa Cruz. In addition to the analytical tests we also made permeability and porosity measurements. Data from all these tests are contained in the attached table.

Study of the analytical data indicates that the results of the ash determinations can be used as a measure of the oil residue in the core samples provided correction is made for carbon dioxide lost by decomposition of carbonates in the cores. Only samples from the 17-ft interval demonstrated appreciable carbonate mineral content. When that carbon content is adjusted for the oxygen which was lost with it, the sum of the carbon dioxide and the carbon and hydrogen is almost equal to the loss obtained in the ash determination.

We have discussed these tests with our analytical group, and the costs for the various tests are as follows:

Ash determination	\$2.50 each in lots of 25 or more
Carbon dioxide by evolution	\$8.00 each in lots of 12 or more
Carbon-hydrogen determination	\$8.00 per test in lots of 12 or more.

To get an effective measure of the carbon-hydrogen residuum in the core samples it appears that we would have to determine carbon dioxide by evolution, and either the carbon-hydrogen content or the ash. Analyses of the costs indicate that the ash determination is preferable to the carbon-hydrogen determination. It is possible that any carbonate minerals present may be restricted to certain strata in the ground and study of additional samples might indicate that it would be unnecessary to determine carbon dioxide by evolution on every core sample studied. We suggest that it would be desirable to obtain enough test information either to confirm or refute this possibility. To accomplish this, we propose that a complete set of cores from a representative core hole in the L-9 area be analyzed both for ash determination and carbon dioxide by evolution. Based upon the results of these tests, a group of cores chosen from appropriate levels in a second representative core hole should be analyzed to establish whether or not the carbonate minerals are restricted to certain intervals.

The content of organic matter in the cores from L-73 seems at first glance to be surprisingly high. We believe, however, that it correlates with low oil recovery from this test pattern. Apparently, only the sample from 17 ft was sufficiently coked to make the residue relatively insoluble in trichloroethane. The reported values for permeability may be of little value because of fractures in the core samples. The high values are undoubtedly the result of such fractures, and probably the permeability of the sample from the 17 ft interval (2760 m.d.) more nearly represents the proper permeability value in the unfractured formation. It is possible that the formation itself is fractured but there is no way to establish whether or not the core samples themselves properly represent this fractured condition.

We have found the results of these preliminary tests quite interesting and believe that we should obtain sufficient information on the L-9 cores to confirm the apparent value of these test data. We shall look forward to your reaction to our proposal.

Very truly yours,

John E. Sherborne

John E. Sherborne, Manager
Production Research Division

RSC:vb
Attachment

cc: R. E. Helander
W. J. Shirley
B. Persson
M. Eurenius

SANTA CRUZ - L-73 CORE SAMPLES

Core Interval Depth, ft	Air Perm. md[1]	Porosity, % by Volume	Extractable Organic Matter, Wt %[2]	Total C - H Det'n Weight, % C	Total C - H Det'n Weight, % H	CO_2 By Evolution, Weight, %	Carbon Loss in CO_2 Wt. %	Ash, Wt. %
17	2,760	19.5	0.5	4.0	0.2[3]	6.2	1.7	91.2
				4.1	0.2			
33	9,000	25.0	7.6	8.5	0.9	<0.3	<0.1	90.1
				8.5	0.9			
45	15,700	17.2	8.7	7.0	0.9	<0.3	<0.1	91.8
				7.2	0.9			

1 Visual examination indicates cores may contain fractures.
2 Extracted with trichloroethane - adjacent samples.
3 Duplicate samples.

COMPARISON OF ORGANIC MATTER CONTENTS DETERMINED BY THE VARIOUS METHODS

Sample	Extractable Matter, Wt. %	Calc. From Ash, Wt. %	Ash Results Corrected for CO_2 Loss, Wt. %	C - H Det'n Corrected for CO_2 Loss, Wt. %
17	0.5	8.8	2.6	2.5
33	7.6	9.9	9.6	9.3
45	8.7	8.2	7.9	7.9

Union Oil Company of California

RESEARCH DEPARTMENT

BREA, CALIFORNIA

March 23, 1959 JES-39

Mr. M. F. Westfall (3)
Husky Oil Company
Cody, Wyoming

Dr. Gosta Salomonsson (3)
Svenska Skifferolje Aktiebolaget
Vastra Gatan 2
Orebro, Sweden

Gentlemen:

We have completed ash and carbon dioxide by evolution determinations on samples from 11 more core holes at Santa Cruz. Although we have additional cores to study and shall complete work on them in the near future, we are making this interim report to permit you to review the data available to date. The data from these tests are contained in the attached tables, and for convenience we have included the results on core holes C-16 and C-19, which were analyzed earlier and discussed in the Engineering Committee Meeting on March 3. We have also included a C-H determination on sample B-5-4 submitted by B. Persson.

In general the data appear to be consistent with our understanding of the process. A rich coked zone exists in the lower intervals of those core holes near a heated well. Apparently oil migrated and gravitated into the hot lower intervals and was subsequently coked therein. If you have any questions regarding these tests or wish additional copies of the test reports, please contact us.

Very truly yours,

John E. Sherborne /RSC

John E. Sherborne, Manager
Production Research Division

RSC:vb
enc.
cc/w: M. Eurenius
R. E. Helander
B. Persson
W. J. Shirley

SWEDISH PROCESS FIELD TEST - SANTA CRUZ
ANALYTICAL TESTS - POST-HEATING CORE SAMPLES
UNION OIL COMPANY OF CALIFORNIA - REPORTED MARCH 24, 1959

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-16			
8-10 FT	7.2	0.4	6.8
10-15	10.5	6.4	4.1
15-20	13.4	8.1	5.3
20-25	11.1	7.8	3.3
25-30	8.5	1.9	6.6
30-35	11.5	0.2	11.3
35-40	13.3	0.4	12.9
40-42	13.0	0.3	12.7
CORE HOLE C-19			
11-15 FT	11.1	9.9	1.2
15-20	13.4	10.0	3.4
20-25	13.5	11.3	2.2
25-30	10.2	6.5	3.7
30-35	8.5	3.2	5.3
35-40	7.7	0.8	6.9
40-44	7.9	0.3	7.6
CORE HOLE C-12			
20-25 FT	8.7	6.1	2.6
25-30	4.9	2.4	2.5
30-35	6.5	0.7	5.8
35-40	11.9	0.3	11.6
40-46	5.9	0.6	5.3

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-16A			
10-15 FT	8.9	6.2	2.7
15-20	13.0	8.1	4.9
20-25	11.6	7.0	4.6
25-30	6.5	1.8	4.7
30-35	11.5	0.2	11.3
35-40	13.9	0.5	13.4
40-42	11.9	0.3	11.6
CORE HOLE C-19A			
15-20 FT	11.9	10.1	1.8
20-25	10.1	8.5	1.5
25-30	7.7	5.7	2.0
30-35	7.4	2.8	4.6
35-40	6.3	0.7	5.6
40-44	6.5	0.4	6.1
CORE HOLE C-17			
11-15 FT	9.3	7.3	2.0
15-20	12.1	10.2	0.9
20-25	10.5	8.7	1.8
25-30	10.1	5.3	4.8
30-35	11.2	1.1	10.1
35-42	11.3	0.6	10.7

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-18			
10-15 FT	11.3	10.7	0.6
15-20	14.0	11.0	3.0
20-25	12.6	9.3	3.3
25-30	9.2	6.6	2.6
30-35	11.6	4.9	6.7
35-40	8.4	1.0	7.4
40-45	6.3	0.2	6.1
CORE HOLE C-20			
15-19 FT	13.9	11.3	2.6
19-20	6.1	3.0	3.1
20-25	13.3	9.5	3.8
25-30	10.9	7.4	3.5
30-35	9.3	5.0	4.3
35-40	9.7	5.3	4.4
40-47	6.4	3.7	2.7
CORE HOLE C-21			
12-15 FT	9.5	6.1	3.4
15-20	9.9	7.3	2.6
20-25	9.5	6.8	2.7
25-30	8.9	3.5	5.4
30-35	8.5	0.4	8.1
35-40	10.7	0.2	10.5
40-47	9.6	0.3	9.3

SWEDISH PROCESS FIELD TEST - SANTA CRUZ
ANALYTICAL TESTS - POST-HEATING CORE SAMPLES
UNION OIL COMPANY OF CALIFORNIA - REPORTED MARCH 24, 1959

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-22			
12-15 FT	8.3	8.9	0
15-20	14.3	6.3	8.0
20-25	6.1	4.4	1.7
25-30	6.0	2.5	3.5
30-35	2.8	1.5	1.3
35-40	8.6	1.7	6.9
40-42	6.2	4.3	1.9
CORE HOLE C-23			
12-15 FT	11.5	10.2	1.3
15-20	10.3	9.5	0.8
20-25	7.5	6.8	0.7
25-30	6.9	5.7	1.2
30-35	11.2	0.7	10.5
35-40	6.2	3.8	2.4
40-42	6.6	0.3	6.3
CORE HOLE C-27			
15-20 FT	9.4	8.1	1.3
20-25	5.9	3.9	2.0
25-30	3.1	0.3	2.8
30-35	4.4	0.2	4.2
35-40	4.7	0.3	4.3
40-43	9.9	2.1	7.8

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-28			
15-20 FT	10.7	8.2	2.5
20-25	6.9	5.0	1.9
25-30	4.2	1.8	2.4
30-35	2.8	0.7	2.1

SAMPLE B-5-4

CARBON CONTENT, % BY WT	59.5
HYDROGEN " " " "	3.2
CO_2 BY EVOL., % BY WT	<0.1

Union Oil Company of California

RESEARCH DEPARTMENT

BREA, CALIFORNIA

April 14, 1959 JES-48

Mr. M. F. Westfall (3)
Husky Oil Company
Cody, Wyoming

Dear Wes:

Our Analytical Laboratory has completed the ash and CO_2 by evolution analyses on the core samples from the Swedish Process Field Test at Santa Cruz. Tabulated data for samples from the last 15 core holes are attached. Other data were reported previously. There are two samples the results for which obviously are or may be anomalous - the 40'-44' interval in C-13 and the 15'-20' interval in C-34. We are obtaining check analyses on these two samples and shall present the data at the Engineering Committee meeting in Santa Cruz on April 23.

We agreed to make the ash determinations for $2.50 per sample and the CO_2 by evolution determination for $8.00 per sample in quantity lots. If because of the quantity of samples processed there should be a savings over these prices, we shall pass these savings on to the project.

If you have any questions regarding these data, we shall be happy to discuss them with you.

Very truly yours,

John E. Sherborne

John E. Sherborne, Manager
Production Research Division

RSC:vb
cc: Dr. Gosta Salomonsson (3)
Mr. M. Eurenius
Dr. R. E. Helander
Mr. B. Persson
Mr. W. J. Shirley

SWEDISH PROCESS FIELD TEST – SANTA CRUZ
ANALYTICAL TESTS – POST-HEATING CORE SAMPLES
UNION OIL COMPANY OF CALIFORNIA – REPORTED APRIL 18, 1959

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-8			
15-20	11.4	7.7	3.7
20-25	8.6	5.7	2.9
25-30	7.7	1.3	6.4
30-35	5.8	0.4	5.4
35-40	7.4	0.6	6.8
40-45.5	6.4	0.1	6.3
CORE HOLE C-13			
10-15	11.3	5.4	5.9
15-20	7.7	5.2	2.5
20-25	7.4	8.8	3.6
25-30	4.0	0.5	3.5
30-35	7.4	0.4	7.0
35-40	10.4	0.9	9.5
40-44	18.2	0.3	17.9
CORE HOLE C-24			
15-20	19.1	6.1	3.0
20-25	8.8	6.8	2.0
25-31	3.9	1.4	2.5
31-37	7.6	0.6	7.0
CORE HOLE C-29			
10-15	11.7	6.8	4.9
15-20	12.4	8.5	3.9
20-25	6.5	4.2	2.8
25-30	5.0	2.6	2.4
30-35	3.0	0.4	2.6
35-40	8.0	0.4	7.6
40-45	7.5	0.5	7.0
CORE HOLE C-32			
5-9	6.2	2.4	3.8
9-15	12.2	10.6	1.6
15-20	10.8	10.7	0.1
20-25	8.3	5.9	2.4
25-30	4.6	1.3	3.3
30-35	3.8	0.6	3.2
35-39	4.5	0.6	3.9

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-9			
14-20	10.4	6.8	3.6
20-25	9.3	5.4	3.9
25-30	6.8	2.0	4.8
30-35	7.5	0.7	6.2
35-40	10.0	0.3	9.7
40-46	9.4	0.6	8.8
CORE HOLE C-14			
10-15	9.9	4.0	5.9
15-20	8.5	4.4	4.1
20-25	6.5	3.8	2.7
25-30	2.4	0.7	1.7
30-35	2.7	0.8	1.9
CORE HOLE C-26			
12-17.5	2.5	0.6	1.9
17.5-20	9.2	6.5	2.7
20-25	8.8	8.2	0.6
25-30	2.4	0.5	1.9
30-35	4.9	0.7	4.2
35-38.5	5.9	1.3	4.6
CORE HOLE C-30			
10-15	11.8	7.2	4.6
15-20	11.2	8.7	2.5
20-25	6.6	4.5	2.1
25-30	8.5	1.4	7.1
30-35	10.0	0.5	9.5
35-40	13.6	0.3	13.3
40-43	8.6	0.4	8.2
CORE HOLE C-33			
6-11	4.2	2.0	2.2
11-15	13.1	10.0	3.1
15-20	10.7	10.7	0.0
20-25	9.4	7.8	1.6
25-30	4.9	1.5	3.4
30-35	5.3	0.5	4.8
35-38	7.4	1.1	6.3
40-49	2.1	0.1	2.0

DEPTH INTERVAL	ASH TOTAL WT LOSS %	CO_2 BY EVOL. WT %	ORGANIC MATTER WT %
CORE HOLE C-11			
20-25	8.2	6.8	1.4
25-30	5.9	2.1	3.8
30-35A	10.5	7.1	3.4
30-35B	3.7	0.3	3.4
35-40	5.3	0.5	4.8
40-45	8.6	0.3	8.3
45-46	9.6	1.6	8.0
CORE HOLE C-15			
9-15	6.5	2.3	4.2
15-20	8.6	2.8	5.8
20-25	7.5	4.3	3.2
25-30	4.1	0.4	4.0
30-35	8.6	2.9	5.7
35-40	10.2	1.3	8.9
40-46	9.2	0.7	8.5
CORE HOLE C-28			
15-20	PREVIOUSLY REPORTED		
20-25	PREVIOUSLY REPORTED		
25-30	PREVIOUSLY REPORTED		
30-35	PREVIOUSLY REPORTED		
35-40	6.0	0.4	5.6
40-46	6.5	0.3	6.2
CORE HOLE C-31			
15-20	12.1	10.6	1.5
20-25	10.0	6.9	3.1
25-30	5.5	3.3	2.2
30-35	6.2	1.1	5.1
35-40	5.2	0.4	4.8
40-50	3.3	0.4	2.9
50-55	5.1	0.4	4.7
CORE HOLE C-34			
10-15	7.4	4.0	3.4
15-20	12.5	15.5	-3.0
20-25	10.4	7.8	2.6
25-30	7.1	1.1	6.0
30-35	10.6	0.5	10.1
35-40	8.7	0.7	8.0
40-42	9.5	0.3	9.2
42-45	5.0	0.1	4.9

SWEDISH [illegible] SS FIELD TEST - S[illegible] A CRUZ

[illegible]ALYTICAL [illegible]S--POST-HEATING CORE SAMPLES

Union Oil Company of California - Analytical Laboratory
March 2, 1959

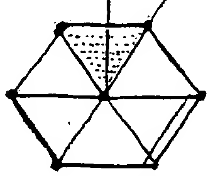
Depth Interval	Ash, Total Wt. Loss	CO_2 by Evol. Wt. %	Organic Matter From Ash, Wt.%	Organic Matter Extracted--Santa Cruz, Wt. %	Organic Matter Extracted - Union, Wt. %	C-H Det'n Corr. C Wt.%	C-H Det'n H %	C-H Atomic Ratio
CORE HOLE C-16								
8-10 ft	7.2	0.4	6.8	-	-	-	-	-
10-15	10.5	6.4	4.1	0.82	3.45	3.1	0.3	$CH_{1.1}$
15-20	13.4	8.1	5.3	0.062	-	-	-	-
20-25	11.1	7.8	3.3	0.027	-	-	-	-
25-30	8.5	1.9	6.6	0.026	-	-	-	-
30-35	11.5	0.2	11.3	0.015	0.87	9.4	0.4	$CH_{0.51}$
35-40	13.3	0.4	12.9	0.038	0.78	11.8	0.5	$CH_{0.51}$
40-42	13.0	0.3	12.7	0.088	0.22	9.8	0.5	$CH_{0.61}$
CORE HOLE C-19								
11-15 ft	11.1	9.9	1.2	1.17	-	-	-	-
15-20	13.4	10.0	3.4	0	-	-	-	-
20-25	13.5	11.3	2.2	0	-	-	-	-
25-30	10.2	6.5	3.7	0	-	-	-	-
30-35	8.5	3.2	5.3	0	-	-	-	-
35-40	7.7	0.8	6.9	0	-	-	-	-
40-44	7.9	0.3	7.6	2.39	4.1	4.5	0.4	$CH_{1.1}$

RSC:vb
3/2/59
Brea

felaktiga värmekonstanter

2. Seven-burner field. 16.3.57. Sal

The burners are assumed to be arranged in the corners and the center of a regular hexagon with an edge length of ~~=~~ 1.21 meter (4 feet). (Corresponds to test L7.) Here the influence of all seven



burners must be considered in ~~each~~ every point of the field. As the arrangement is symmetric ~~al~~ only one ~~sixth~~ twelfth of the whole hexagon needs to be considered.

~~St~~ The equation for the heat distribution around one burner is:

$$\Delta t = \frac{H}{4\pi \cdot \lambda} \cdot \varphi\left(\frac{s \cdot g}{4\lambda} \cdot \frac{r^2}{\tau}\right); \qquad (1)$$

or:

$$\Delta t = 73.8 \cdot \varphi\left(31.4 \cdot \frac{r^2}{\tau}\right) \qquad (2)$$

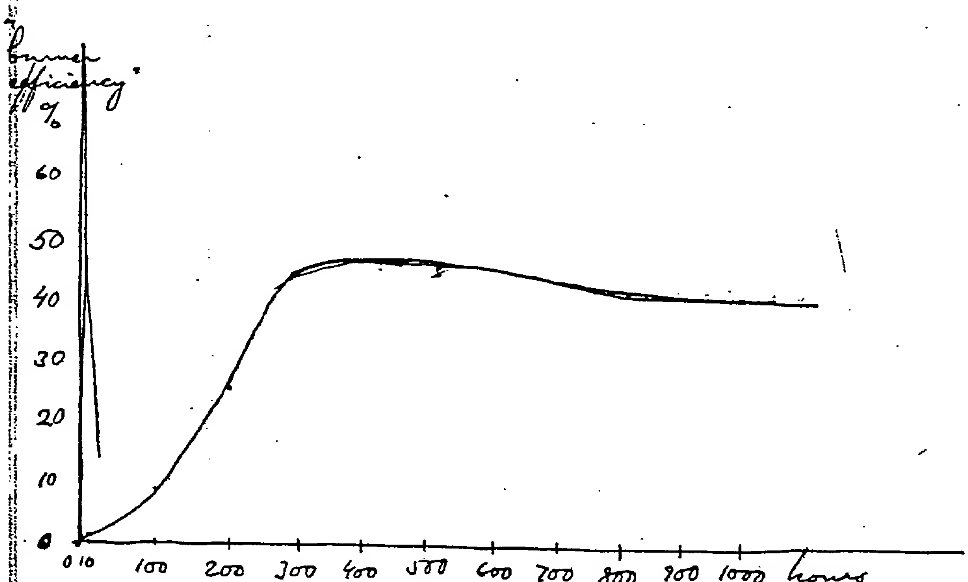
The temperatures are calculated for a number of times, viz. τ = 10, 100, 200, ~~400~~ 300, 500, 800, 1000, ~~=~~ 2000, 4000 and 10.000 hours.

As 300 °C is the lowest temperature that is of any interest in this connection, and since all actual temperatures are the sum of seven increments, no Δt temperatures smaller than $\frac{300}{7} \doteq 44$ °C need to be considered, that is no $\frac{r^2}{\tau}$-values smaller than 0.0150, corresponding to r = 750 cm = 7.5 meters.

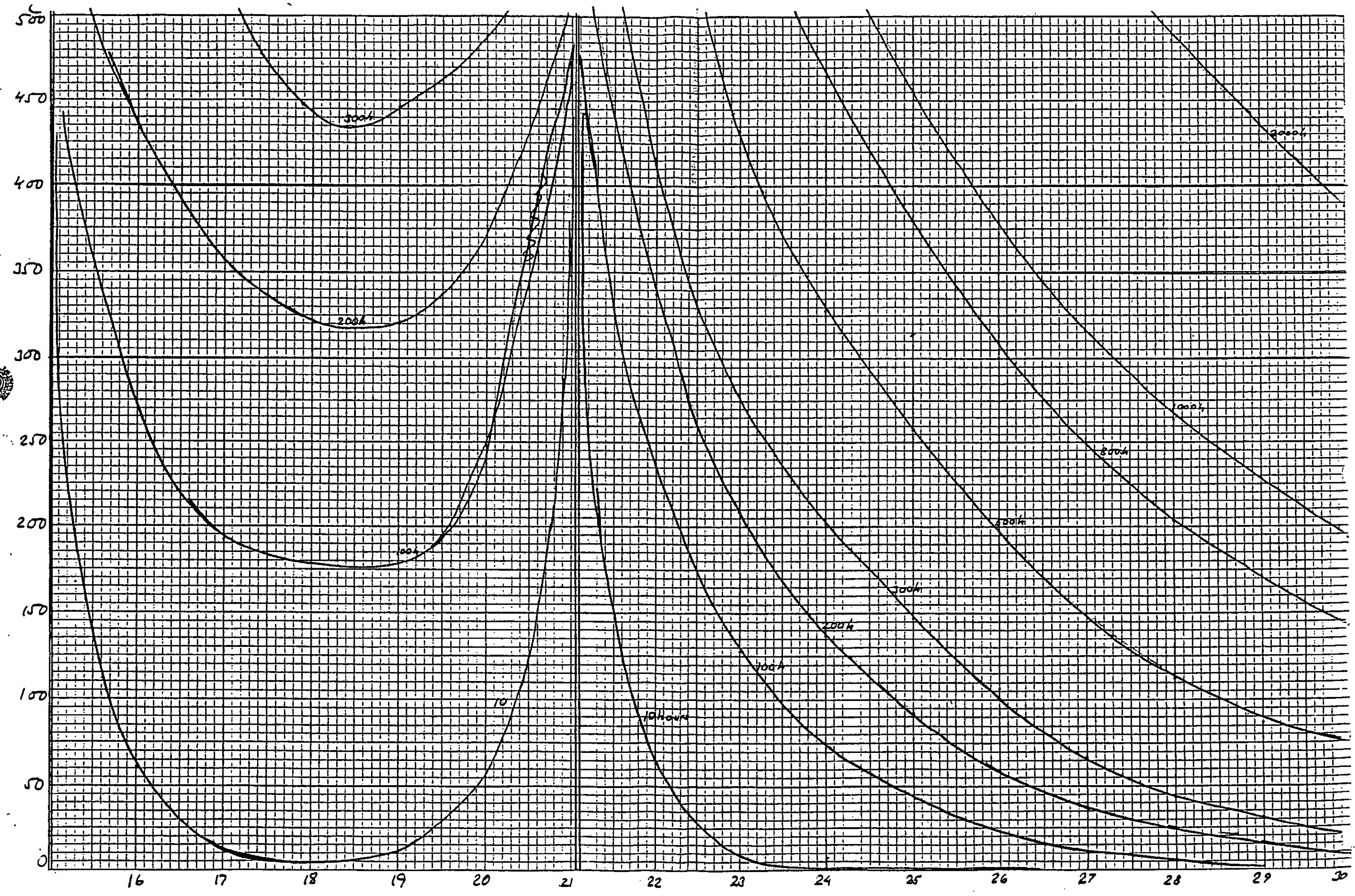
According to the graphs, page = , the pyrolyzed amount of rock per one ~~meter~~ cm burner length is:

$\boxed{\Delta H = 10\,kcal/h,cm}$ cm^3/cm burner length (hole distance = 4 ft)

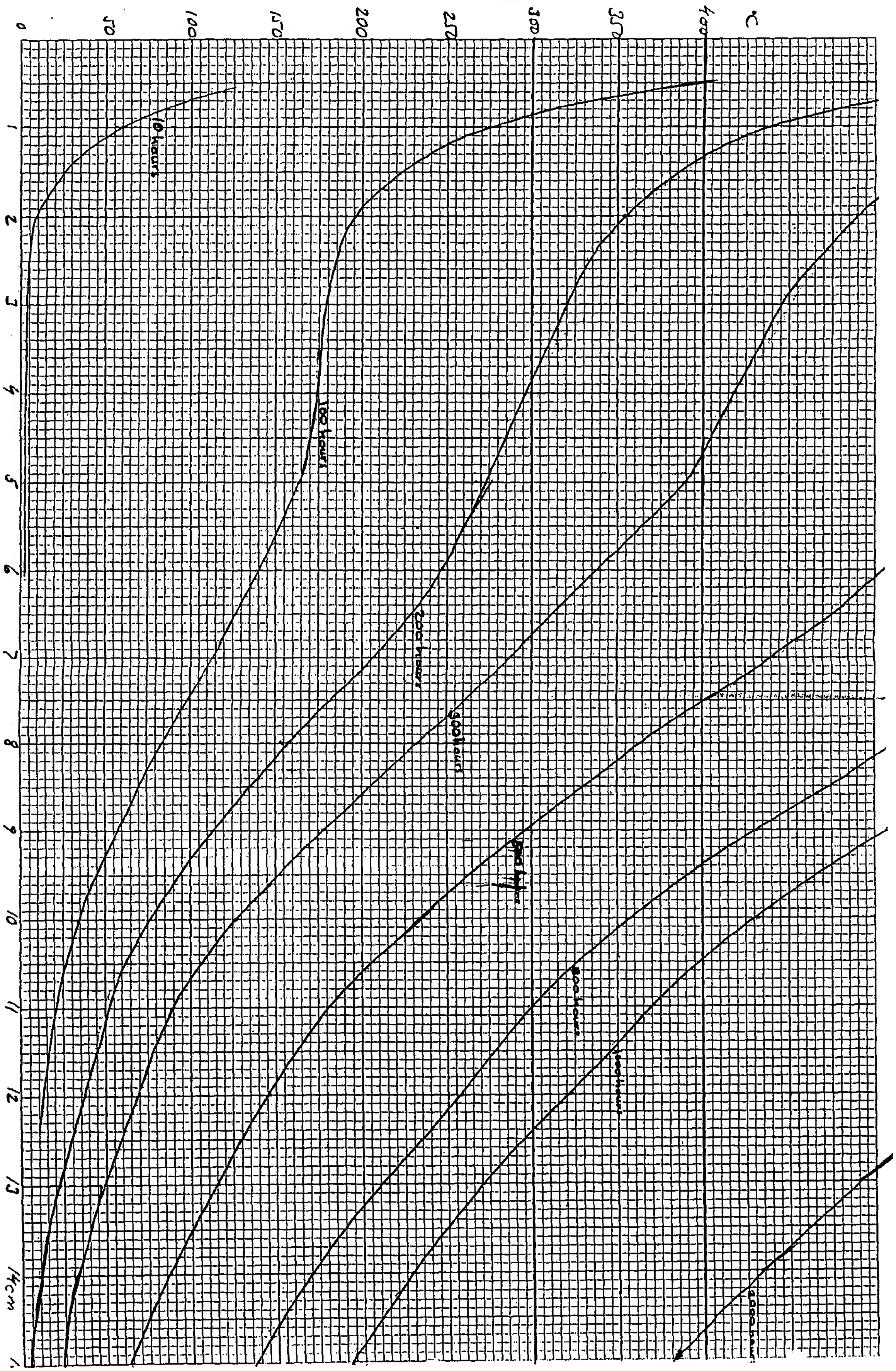
time hours	$\Sigma H_{kcal/cm}$	> 400°C (A)	350–400°C (B)	300–350°C (C)	Σ (A) + 0.75(B) + 0.3(C)
10	7·100	56 cm^3 =		56·0.167 = 9.35 kcal utilized heat	= 13%
100	7·1000	3500		585	8.4%
200	7·2000	21.920		3670	26%
300	7·3000	54.760		9160	44%
500	7·5000	94.200		15.800	45%
800	7·8000	137.000		22.900	41%
1000	7·10.000	168.600		28.200	40%



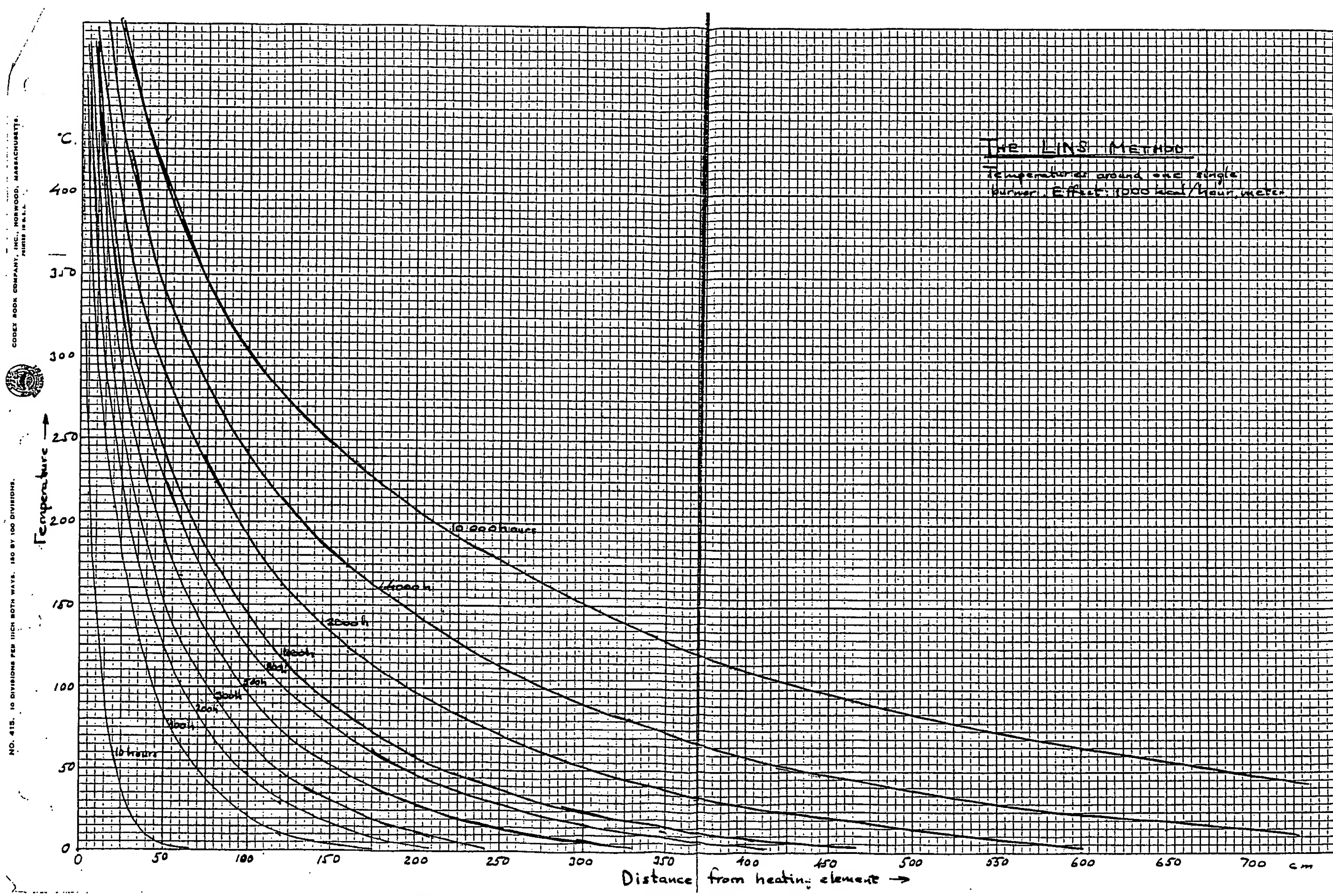
For another heat input, f.i. 20,000 BTU/h,ft = 11 kcal/h, cm, we find from the logarithmic $x/\varphi(x)$-diagram that, if $\varphi(x)$ has to be lower in the ratio $\frac{10}{11} = 0.91$, $x \left(= \frac{r^2}{t}\right)$ has to be increased about 12%, meaning that t has to be decreased 12% to make the figures approximately valid. Thus the efficiency 26% is valid at $\frac{200}{1.12} = \sim 178$ hours instead of 200 h.



NO. 415. 10 DIVISIONS PER INCH BOTH WAYS. 150 BY 100 DIVISIONS.
CODEX BOOK COMPANY, INC., NORWOOD, MASSACHUSETTS.
PRINTED IN U.S.A.



0
50
100
150
200
250
300
350
400
°C
1
2
3
4
5
6
7
8
9
10
11
12
13
14cm
10 hours
100 hours
200 hours
300 hours
800 hours


THE LINS METHOD
Temperatures around one single burner. Effect: 1000 kcal/hour, meter
°C
Temperature →
400
350
300
250
200
150
100
50
0
10 000 hours
4000 h
2000 h
1000 h
800 h
500 h
300 h
200 h
100 h
10 hours
0
50
100
150
200
250
300
350
400
450
500
550
600
650
700
cm
Distance from heating element →

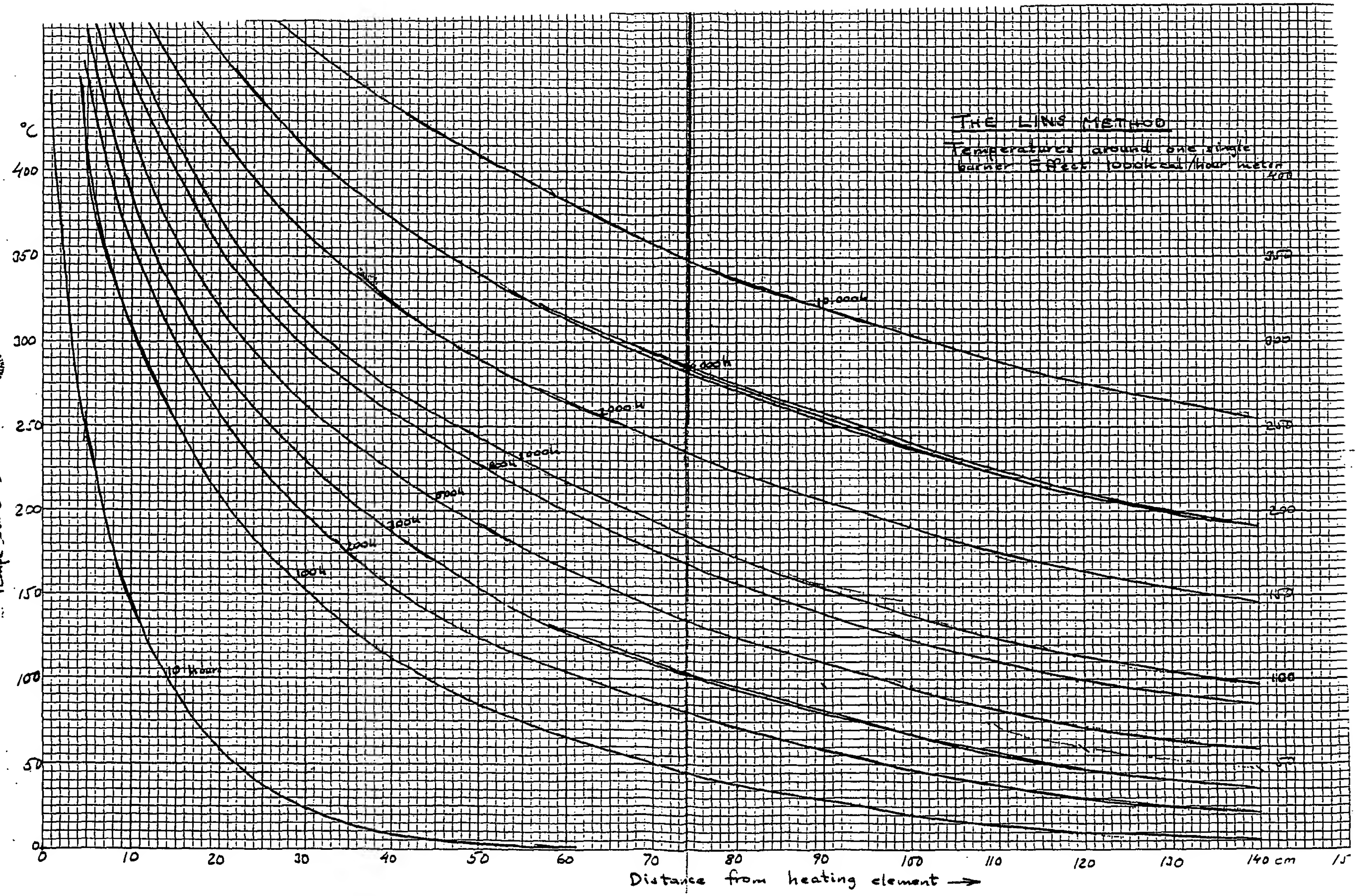

THE LINS METHOD
Temperatures around one single burner Effect 1000kcal/hour meter
°C
400
350
300
250
200
150
100
50
0
Temperature →
0
10
20
30
40
50
60
70
80
90
100
110
120
130
140 cm
Distance from heating element →
10 hours
100h
200h
300h
500h
800h
1000h
2000h
5000h
10.000h

Table of [illegible]

r cm \ T hours	10	100	200	300	500	800	1000	2000	4000	10.000
5	243	408	458	482						
10	144	306	356	387	424	458	472			
20	58.2	208	257	285	322	356	374	423	472	
30	24.4	153	198	231	263	297	313	364	415	479
40	8.3	111	158	187	224	257	271	322	373	441
50	2.6	85	124	157	192	227	244	290	338	408
60	1.0	64	103	126	167	199	216	264	314	382
80	–	37	71	93	123	158	174	224	272	336
100		21	47	68	97	124	145	191	243	306
0		9.0	29	46	72	98	113	162	209	273
140		5.8	21	35	59	85	97	145	192	257
160			13	26	47	71	83	121	174	241
180			8.1	18	36	58	70	110	160	224
200				12	28	48	59	97	145	207
250					14	29	38	72	112	176
300					7	17	24	52	90	152
350						9	15	38	74	129
400							8	28	58	111
450								21	47	97
500								14	38	85
0									24	64
700									15	49

Table of [illegible]

3.30	5.53	6.21	6.54						
1.96	4.15	4.83	5.25	5.75	6.21	6.40			
0.79	2.82	3.48	3.86	4.37	4.83	5.08	5.74	6.40	
0.33	2.07	2.69	3.13	3.57	4.02	4.25	4.94	5.63	6.50
0.113	1.50	2.14	2.54	3.04	3.48	3.68	4.37	5.07	5.98
0.0357	1.16	1.68	2.13	2.60	3.08	3.31	3.93	4.59	5.53
0.013	0.87	1.40	1.71	2.26	2.70	2.92	3.58	4.25	5.19
0.002	0.50	0.96	1.26	1.66	2.14	2.36	3.04	3.68	4.56
	0.28	0.64	0.92	1.31	1.68	1.96	2.57	3.30	4.15
	0.122	0.395	0.62	0.98	1.33	1.525	2.19	2.84	3.70
	0.078	0.28	0.475	0.795	1.16	1.31	1.96	2.60	3.49
	0.0033	0.173	0.353	0.630	0.96	1.13	1.64	2.36	3.27
	0.0016	0.110	0.250	0.490	0.78	0.945	1.49	2.16	3.03
	0.0009	0.0077	0.160	0.380	0.645	0.795	1.31	1.96	2.81
		0.00185	0.0068	0.19	0.390	0.570	0.975	1.62	2.38
		0.00055	0.0021	0.094	0.232	0.33	0.705	1.22	2.06
			0.0008	0.0375	0.126	0.20	0.52	1.00	1.75
				0.017	0.077	0.113	0.38	0.79	1.50
				0.009	0.037	0.079	0.286	0.635	1.32
					0.078	0.036	0.19	0.57	1.16
					0.0055	0.012	0.095	0.33	0.87
							0.038	0.20	0.66

Table [illegible]

	T = hours	10	100	200	300	500	800	1000	2000	4000	10.000
	T = sec $\cdot 10^{-6}$	0.036	0.360	0.720	1.080	1.800	2.88	3.60	7.20	14.40	36.00
	$\frac{31.4}{T_{sec}} \cdot 10^{+6}$	872	87.2	43.6	29.1	17.44	10.90	8.72	4.36	2.18	0.872
r	B $\frac{r^2}{cm^2} \cdot 10^{-3}$										
5	0.025	21.8	2.18	1.09	0.728	0.437	0.273	0.218	0.109	0.0545	0.0218
10	0.1	87.2	8.72	4.36	2.91	1.74	1.09	0.872	0.436	0.218	0.0872
20	0.4	349	34.9	17.45	11.83	6.98	4.36	3.49	1.745	0.873	0.349
30	0.9	786	78.6	39.3	26.2	15.72	9.83	7.86	3.93	1.965	0.786
40	1.6	1400	140	70	46.7	28.0	17.5	14.0	7.00	3.50	1.40
50	2.5	2180	218	109	72.8	43.7	27.3	21.8	10.9	5.45	2.18
60	3.6	3140	314	157	104.7	62.8	39.3	31.4	15.7	7.85	3.14
80	6.4	5600	560	280	187.7	112.0	70.0	56.0	28.0	14.0	5.60
00	10.0	8720	872	436	291	174.4	109.0	87.2	43.6	21.8	8.72
	14.4	13.580	1358	679	452,7	271,6	169,8	135,8	67.9	33,95	13,58
140	19.6	17.400	1740	870	580	348.0	217.5	174	87	43.5	17.40
160	25.6	22300	2230	1150	743.3	446	279	223	115	57.5	22.3
180	32.4	28.200	2820	1410	940	564	352,5	282	141	70,5	28,2
200	40.0	34.900	3490	1745	1183	698	436	349	174,5	87,3	34,9
250	62.5	54.600	5460	2730	1820	1092	682,5	546	273	136,5	54.6
300	90.0	78.600	7860	3930	2620	1572	983	786	393	196,5	78,6
350	122.5	107.000	10.700	5.350	3.567	2.140	1338	1070	535	267,5	107
400	160.0	140.000	14.000	7.000	4670	2800	1750	1400	700	350	140
450	202.5	173.000	17.300	8650	5767	3460	2162,5	1730	865	432,5	173
500	250.0	218.000	21800	10900	7280	4370	2730	2180	1090	545	218
6	360.0	314.000	31.400	15.700	10.470	6280	3930	3140	1570	785	314
700	490.0	428.000	42.800	21.400	14.267	8560	5350	4280	2140	1070	428

2	8	1	–	–	–	–	1	10
3	1	1	–	–	–	–	1	3
4	1	1	–	–	–	–	1	3
5	1	1	–	–	–	–	1	3
6	–	1	–	–	–	–	1	2
7	–	1	–	–	–	–	1	2
8	–	1	–	–	–	–	1	2
9	–	1	–	–	–	–	1	2
10	–	–	–	–	–	–	–	1
11	–	–	–	–	–	–	–	–
12	–	–	–	–	–	–	–	–
13	–	–	–	–	–	–	–	–
14	–	–	–	–	–	–	–	–
15	–	–	–	–	–	–	–	–
16	58	1	–	–	–	1	1	61
17	8	1	–	–	–	1	1	11
18	1	1	–	–	–	1	1	4
19	1	1	–	–	–	1	8	11
20	1	1	–	–	–	1	53	56
21	–	–	–	–	–	–	425	425
22	–	–	–	–	–	–	64	64
23	–	–	–	–	–	–	9	9
24	–	–	–	–	–	–	1	1
25	–	–	–	–	–	–	1	1
26	–	–	–	–	–	–	1	1
27	–	–	–	–	–	–	–	–
28	–	–	–	–	–	–	–	–
29	–	–	–	–	–	–	–	–
[illegible]	–	–	–	–	–	–	–	–

1	208	17	10	6	6	10	17	274
2	112	30	8	5	5	8	30	195
3								
4								
5	21	63	5	3	3	5	63	163
6								
7	6	50	3	1	1	3	50	114
8								
9								
10	2	13	2	1	1	2	13	34
11								
12								
13								
14	1	3	1	—	—	1	3	9
15								
16	208	13	7	6	7	13	19	273
17	112	15	5	2	5	15	37	191
18	75	17	2	1	2	17	64	178
19	38	15	1	—	1	15	108	178
20	20	13	—	—	—	13	203	249
21	10	10	—	—	—	10	(440)	470
22	6	7	—	—	—	7	217	237
23	2	5	—	—	—	5	115	127
24	1	2	—	—	—	2	67	72
25	—	1	—	—	—	1	40	42
26	—	—	—	—	—	—	22	22
27	—	—	—	—	—	—	10	10
28							6	6
29							2	2
30							1	1

					—	[illegible]	[illegible]	[illegible]
2	155	60	25	14	14	25	60	353
3	103	77	23	8	8	23	77	319
4	72	93	19	4	4	19	93	304
5	47	102	7	1	1	7	102	267
6	30	98	10	1	1	10	98	248
7	21	87	6	—	—	6	87	207
8	13	69	3	—	—	3	69	157
9	7	57	2	—	—	2	57	113
10	3	35	1	—	—	1	35	75
11	2	24	—	—	—	—	24	50
12	1	18	—	—	—	—	18	37
13	—	11	—	—	—	—	11	22
14	—	5	—	—	—	—	5	10
15	—	2	—	—	—	—	2	4
16	257	34	23	21	23	34	45	437
17	155	40	19	13	19	40	70	356
18	103	42	13	7	13	42	102	322
19	72	40	8	2	8	40	150	320
20	47	35	4	1	4	35	250	376
21	30	28	2	1	2	28	>600	>670
22	22	24	1	—	1	24	265	337
23	13	18	1	—	1	18	158	209
24	7	13	—	—	—	13	105	138
25	3	8	—	—	—	8	73	92
26	2	4	—	—	—	4	48	58
27	1	2	—	—	—	2	31	36
28	—	1	—	—	—	1	23	25
29	—	1	—	—	—	1	14	16
30	—	—	—	—	—	—	7	7

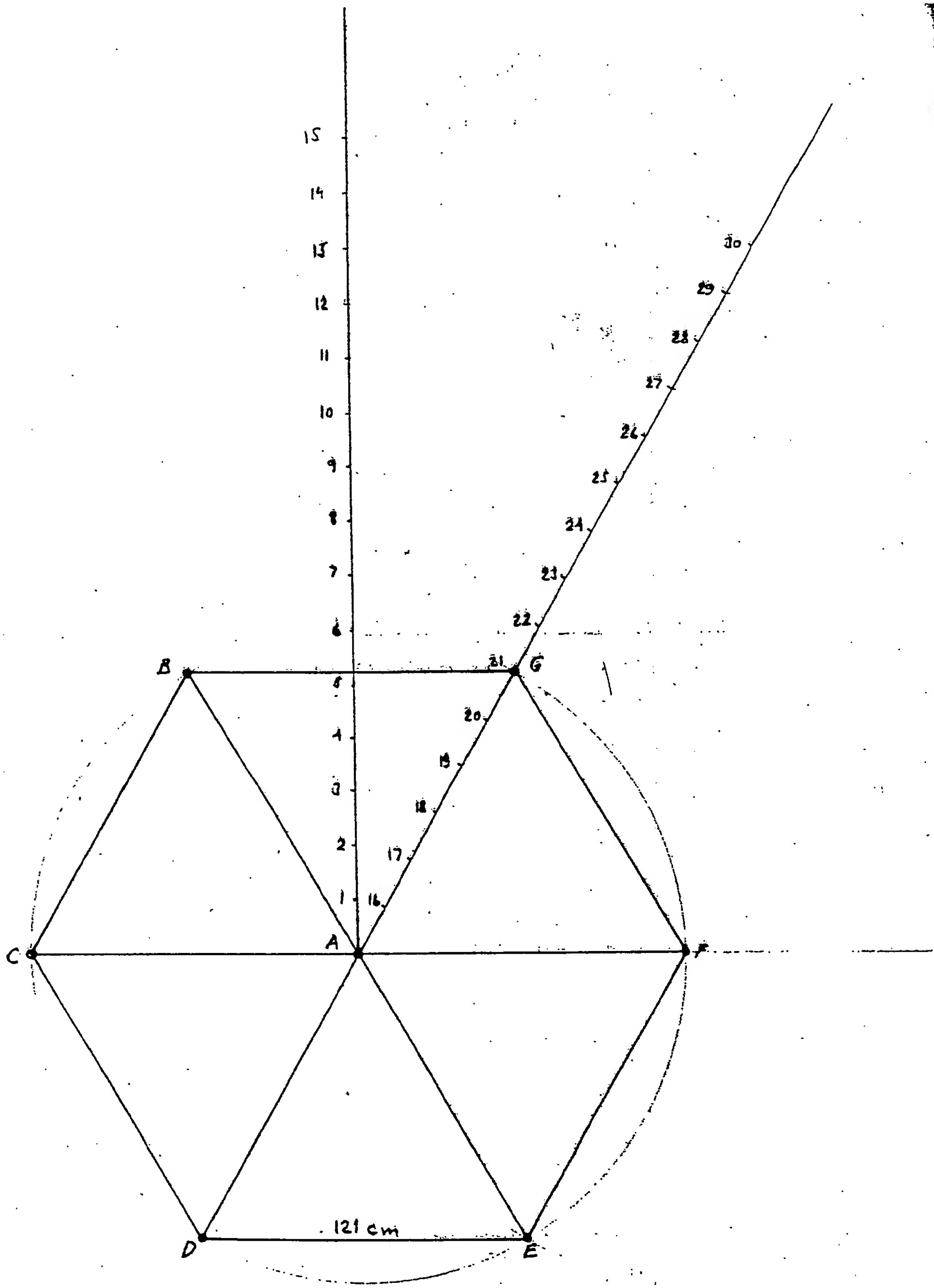
1	286	63	45	36	36	45	63	574
2	188	81	42	28	28	42	81	490
3	127	100	39	20	20	39	100	445
4	94	115	33	14	14	33	115	418
5	67	125	28	8	8	28	125	389
6	47	120	22	3	3	22	120	337
7	36	109	16	1	1	16	109	288
8	26	91	12	–	–	12	91	232
9	18	73	7	–	–	7	73	178
10	12	54	3	–	–	3	54	126
11	7	40	1	–	–	1	40	89
12	2	33	–	–	–	–	33	68
13	1	23	–	–	–	–	23	47
14	–	16	–	–	–	–	16	32
15	–	12	–	–	–	–	12	24
16	286	53	39	36	39	53	66	572
17	188	59	33	26	33	59	93	591
18	127	61	26	17	26	61	125	443
19	94	59	20	12	20	59	184	448
20	67	53	15	7	15	53	279	489
21	47	45	9	2	9	45	>700	>800
22	36	38	5	1	5	38	293	416
23	26	33	2	–	2	33	190	276
24	18	26	1	–	1	26	129	201
25	12	20	–	–	–	20	95	147
26	7	15	–	–	–	15	69	106
27	2	9	–	–	–	9	47	67
28	1	3	–	–	–	3	37	44
29	–	2	–	–	–	2	27	31
30	–	1	–	–	–	1	18	20

2	224	111	66	49	49	66	111	676
3	165	131	61	40	40	61	131	629
4	124	151	55	32	32	55	151	600
5	94	163	49	24	24	49	163	566
6	72	158	42	17	17	42	158	506
7	57	143	35	13	13	35	143	441
8	47	121	28	10	10	28	121	365
9	37	100	23	7	7	23	100	297
10	29	79	18	5	5	18	79	233
11	22	64	13	2	2	13	64	180
12	16	54	10	1	1	10	54	146
13	12	44	7	1	1	7	44	116
14	8	35	5	—	—	5	35	88
15	6	27	2	—	—	2	27	64
16								
17								
18	165	88	47	36	47	88	163	634
19	124	85	40	28	40	85	220	622
20	94	79	33	21	33	79	313	652
21								
22	59	62	19	12	19	62	327	560
23	47	55	15	8	15	55	227	432
24	37	47	12	6	12	47	167	328
25	29	40	8	3	8	40	126	254 ~~314~~
26	22	33 ~~16~~	6	2	6	33 ~~16~~	96	198
27	16	25	3	1	3	25	72	145
28	12	18	2	—	2	18	60	112
29	8	15	1	—	1	15	47	87
30	6	12	—	—	—	12	37	77

1								
2								
3								>500
4								
5								
6								
7	84	177	55	28	28	55	177	604
8	70	152	47	23	23	47	152	574
9	57	128	40	19	19	40	128	431
10	47	107	34	14	14	34	107	357
11	38	91	28	11	11	28	91	298
12	32	80	24	9	9	24	80	258
13	26	67	20	7	7	20	67	214
14	21	55	15	5	5	15	55	171
15	17	45	12	3	3	12	45	137
16								
17								
18								
19								>500
20								
21								
22								
23	70	80	30	21	30	80	261	572
24	57	70	25	16	25	70	202	465
25	47	60	21	13	21	60	158	380
26	38	52	17	10	17	52	124	310
27	32	43	13	8	13	43	100	252
28	26	34	10	5	10	34	85	204
29	21	30	8	3	8	30	71	171
30	17	26	6	2	6	26	58	141

1								
2								
3								
4								>500
5								
6								
7								
8								
9	70	144	50	26	26	50	144	510
10	58	121	44	22	22	44	121	432
11	48	105	37	19	19	37	105	370
12	42	93	32	15	15	32	93	322
13	35	79	27	12	12	27	79	271
14	30	66	23	10	10	23	66	228
15	25	56	20	8	8	20	56	193
16								
17								
18								
19								
20								>500
21								
22								
23								
24	70	83	34	24	34	83	220	548
25	58	73	28	20	28	73	175	455
26	48	63	24	17	24	63	140	379
27	42	53	20	14	20	53	114	316
28	35	44	17	10	17	44	97	264
29	30	40	14	8	14	40	83	229
30	25	34	12	6	12	34	70	193

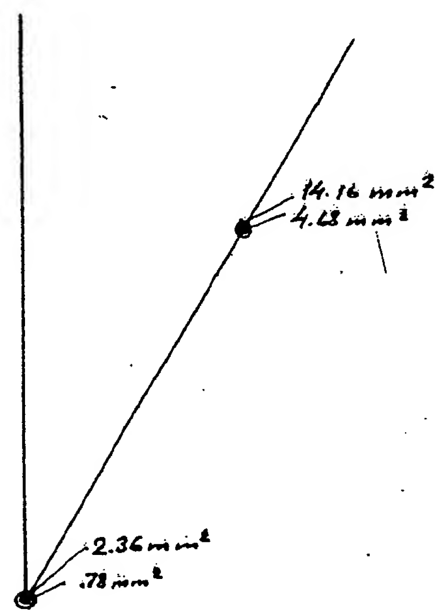
								>500
10	97	173	79	49	49	79	173	699
11	86	155	71	43	43	71	155	624
12	76	137	65	38	38	65	137	576
13	68	120	57	34	34	57	120	490
14	59	107	50	30	30	50	107	433
15	52	95	44	26	26	44	95	382
								>500
25	97	113	59	45	59	113	225	711
26	86	103	52	40	52	103	192	628
27	76	92	46	36	46	92	165	553
28	68	80	41	32	41	80	144	486
29	59	75	37	28	37	75	125	436
30	52	67	33	25	33	67	111	388



15
14
13
12
11
10
9
8
7
6
5
4
3
2
1
16
17
18
19
20
21
22
23
24
25
26
27
28
29
30
A
B
C
D
E
F
G
121 cm

Pyrolyzed area after 10 hours.

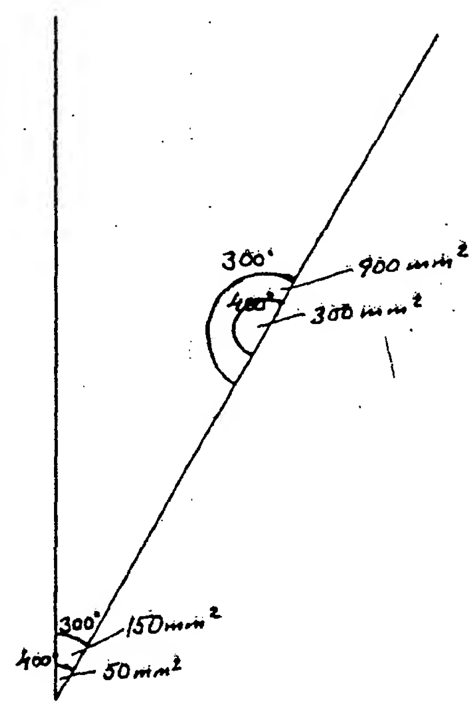
$14\ mm^2$, corresponding to $14 \cdot 4 = 56\ cm^3$ rock per cm of burner length.



All figures refer to whole hexagonal pattern though the drawing shows only $\frac{1}{12}$.

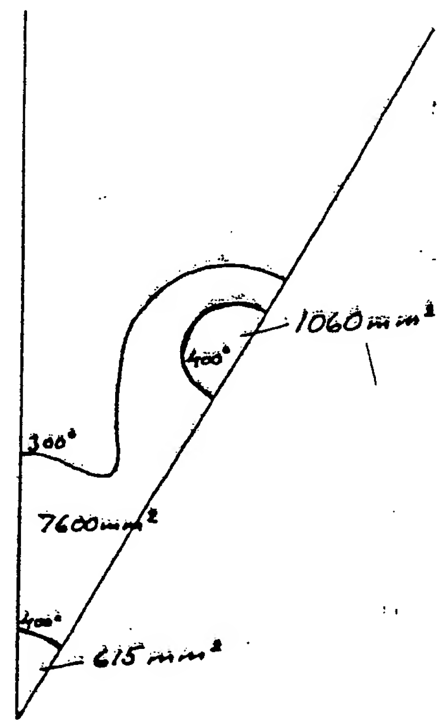
Pyrolyzed area after 100 hours.

$875\ mm^2 = 3500\ cm^3/cm$ burner length.



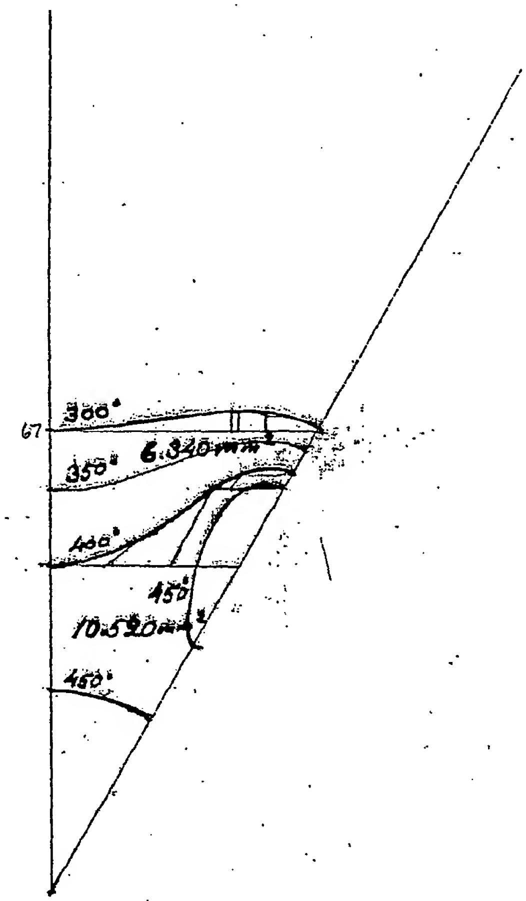
Pyrolyzed area after 200 hours.

$5480\ mm^2 = 21.920\ cm^3/cm$ burner length



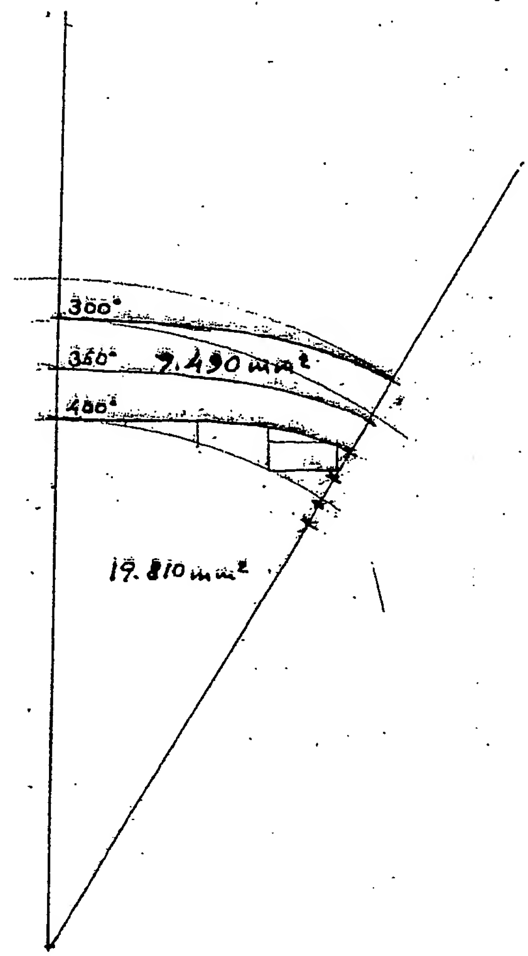
Pyrolyzed area after 300 hours.

13.690 mm² = 54.760 cm³/cm burner length



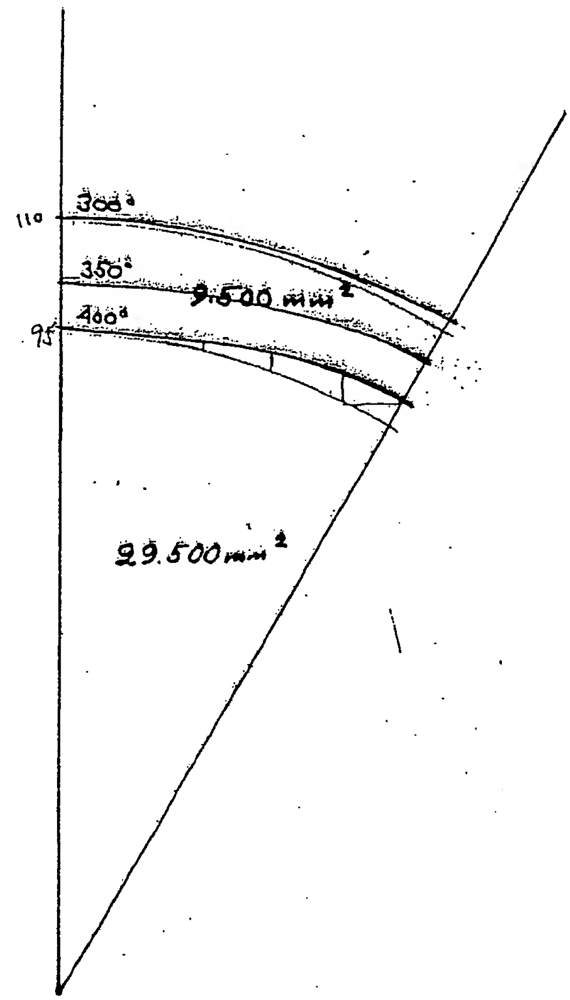
Pyrolyzed area after 500 hours:

23.550 mm² = 94.200 cm³/cm burner length



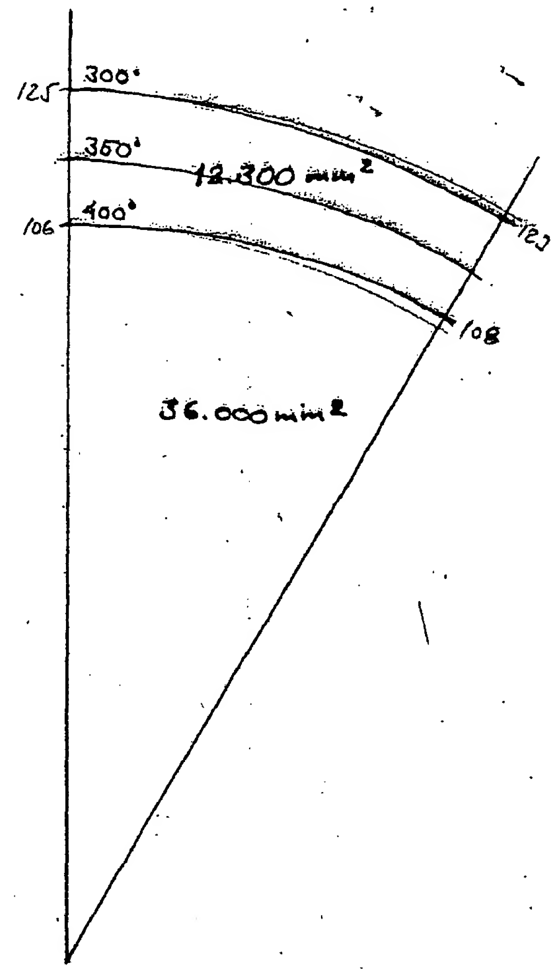
Pyrolyzed area after 800 hours.

$34.250\ mm^2 = 137.000\ cm^3/cm$ burner length.



Pyrolyzed area after 1000 hours

42.150 mm² = 168.600 cm³/cm burner length



OIL RECOVERY FROM TAR SAND

WITH THE LINS METHOD

Report on field tests at SANTA CRUZ, CALIFORNIA

1955 - 1957

conducted by Husky Oil Company

and Svenska Skifferolje Aktiebolaget

SUMMARY

The tar in the tar sand can be transformed to gas, oil and a carbonaceous residue if heated to about 750°F. The objectives of the test work carried out at Santa Cruz during March 1955 through December 1957 and described in this report, were:

- To develop a gas-fired burner, suitable for commercial-scale heating in-situ of a tar sand formation, and
- To study the heat transfer and the flow of produced fluids in the formation.

Starting from a preliminary burner design, developed in the laboratories of Svenska Skifferolje Aktiebolaget, a number of single-burner tests were performed in vertical bore-holes in the tar sand formation. It was found that the most important problem in the burner design was to make long enough burners while maintaining an even heat distribution over the entire length of the burner. Local heat concentrations would tend to damage the burner or the casing.

The original burner would heat only a layer, less than 10 ft thick. The following tests resulted in improvement in the original design. By re-circulating a certain amount of exhaust gas within the burner and by shielding the hottest part of the burner with a concentric steel tube, a small improvement in heat distribution was obtained.

Even this was, however, insufficient for the heating of the useful portion of the tar sand formation at the test site, which was 35 to 40 feet. Therefore, in the first larger-scale test - including 100 burners and covering 4750 sq ft where oil recovery data were to be studied besides the performance of the burners, an attempt was made to obtain the desired 40-foot heat distribution by moving the burners up and down in the wells at regular intervals (every 4 or 8 hours). This method gave a fairly good overall heat distribution, but the momentary, local heat concentrations caused repeated failures in the burner casings, which were made of carbon steel. A number of casings were replaced with new alloy-casings (2.5, 5, 9 and 25 % chromium-

steel alloys) and the test was continued on a reduced scale.

Concurrent with this test the work on improved burner construction continued and resulted in the so-called sand burner. In this burner a fluidized bed of sand is used for the distribution of heat. No exhaust gas recirculation is used and no thermal shields are necessary. Already in the first series of preliminary tests with sandburners heated intervals of up to 34 feet were obtained. Due to the better distribution of the heat along the whole length of the burner it is anticipated that the burner can be manufactured of less expensive construction materials than those used in the original burners and in the sand burners tested so far, where 25/12 chromium-nickel steels were used in the hottest parts and 18/8 steel in the adjacent parts.

As soon as the superior effect of the sand burner was proven, the remaining burners in the earlier 100-hole test were replaced with sand burners.

Preliminary data on heat transfer in the formation were calculated from the temperature observations. These data show that reasonable heat transfer rates can be achieved in tar sand.

In order not to waste heat on water vaporization as much as possible of the ground water present should be removed from the deposit.

No quantitative recovery data were obtained from the 100-burner test because of the irregular operation. The nature of the produced oil was aromatic. Most of the oil was rather light, with gravities between 20 and 35° API, and lightcolored but unstable.

Samples of the produced gas had hydrogen sulfide contents up to 12 % and heat values of 800 to 1000 BTU/st cuft. In one test it was found that sweetened, produced gas is a suitable burner fuel. In all other tests propane was used as fuel.

Current tests, not described in this report, include a new 100-hole test, covering an area of about 7400 sq ft, operated with sand burners and especially intended to give information about obtainable oil and gas yields. Further the first part of a systematic study of the different factors, influencing the efficiency of sand burners and a series of laboratory studies on the relations between heating rates, reaction temperatures, product yields, etc. are being conducted.

Problems, requiring further research work, include:

1. Studies of burner construction materials.
2. Studies of oil and gas recoveries.
3. Development of longer burners.

TAR SAND TESTS AT SANTA CRUZ, CALIFORNIA -

February 1955 to December 1957

INTRODUCTION

During 1953 and 1954 some preliminary studies were made by the Research Department of Svenska Skifferolje Aktiebolaget on the use of the Ljungstrom In Situ Method (LINS Method) for oil recovery from tar sand deposits. The work included analyses of a few samples of tar sand (taken from outcroppings in California and Alberta), some model scale studies on artificial mixtures of sand and tar and some preliminary work on a gas burner to be used instead of the electrical heater, used in the commercial Ljungstrom field in Sweden.

It was found, that only results of very limited value and applicability could be obtained in this way. The tar sand was found to be very nonuniform in physical and chemical properties and small laboratory samples could not yield enough information. Heat transfer, heater design, flow of gases and liquids, and obtainable product yields would depend on a plurality of field factors, which could not be duplicated in the laboratory. It was thus decided that further research on this project should be concentrated to studies in an actual tar sand field.

A tar sand deposit, located between Laguna Creek and Majors Creek, about 9 miles northwest of Santa Cruz, California, was considered a suitable area for the field tests. After core drilling in the area in March and April 1955, a test site was chosen about 500 ft southwest of the Calrock Quarry. A number of single-burner tests were started here during the summer and fall of 1955 for the purpose of studies of burner performance and heat and product flow in the formation.

After several tests had been started in this area, new tests were begun in a new area, north of the quarry. Later, in May 1956, all testing equipment was moved to this area and all subsequent testing has been done there, including a number of single-burner tests and three sevenburner tests. Besides the above-mentioned purposes, the purpose of the sevenburner tests was to obtain sufficient quantities of the produced oil and gas to permit reliable analyses to be made.

Finally, in July 1956, a hundredburner test was started with the objective of obtaining operation and yield data which would be necessary for an evaluation of the commercial possibilities of the LINS Method. This report refers to all field tests up to and including this first hundred burner test. A new hundred burner test was started in February 1958, and is still in operation.

Descriptions of the general test arrangements are given below and detailed descriptions and discussions of the individual tests follow in the Appendix.

GENERAL DESCRIPTION OF THE TESTS

Description of the deposit

When the first test site was chosen, it was felt (based on experiences from the Swedish Ljungstrom field) that the tar sand deposit should be covered by at least 25 ft of overburden (soil, limestone, shale etc.) in order to ensure a gas-tight seal over the pyrolyzed area. This condition was met at the chosen location, where the average overburden thickness was about 55 ft (mainly shale) and the average tar sand layer thickness was about 45 ft. The tar content of this layer was between 6 and 12 % by weight. The tar sand also contained some streaks of clay. Core analyses are included in the test descriptions in the Appendix. The tests L2, 21, 22, 3, 31, 4, 4A, 41, 42, 5, 51, 52, 100, 101, 102, and 103 were located in this area.

The second test site was chosen in order to study the possibilities of the utilization of tar sand deposits without overburden, i.e., if the leakage of products through the surface could be kept within reasonable limits. Here the tar sand was covered by only 7 - 10 ft of soil. Above 45 feet the tar sand was fairly uniform containing 8 to 15 % by weight of tar. Below this level the tar sand was lean and less uniform.

Burner and gas wells

Most of the burner wells were drilled with 4 3/4 inch rock bits to various depths between 40 and 85 ft. Burner casings in most tests consisted of standard 2½-inch pipe (of carbon steel or chromium-alloy steel), closed at its lower end and with its upper end extending ½ ft above the ground in most cases.

The gas wells were of two kinds: concentric wells around the burner casings, and separate wells drilled some distance from the burner well. In a concentric well, a larger gas casing (usually 4-inch pipe) was set around the burner casing. This gas casing penetrated a few feet into the tar sand layer and the fluids were thus produced through the annulus between the casings. The gas well casing was cemented against the rock above its open, lower end. Above the cement the annulus was filled with sand. The upper end of the gas well casing was sealed against the extending end of the burner well casing (by welding or a bushing type connection) and had a side outlet through which the produced fluids were withdrawn. These fluids were conducted through production lines to condensing and separating equipment.

The separate gas wells were drilled through the tar sand interval with a 3 3/4-inch bit. The casings extended to the top of the tar sand in some wells and in others a slotted casing was run to the bottom of the well.

Burners

The original burner, named Type A, consisted of a narrow pipe of varying length for the supply of fuel-air mixture, a conical enlargement which acted as a flameholder, and a 1 inch (in some of the earlier tests 3/4-inch) burner tube, of a length varying in different tests from 5 to 35 ft. The function of the burner tube was to conduct the hot combustion gases to the bottom of the

Row	Hours of oper. at time	Heat input in 10^6 BTU in burner no. 1	2	3	4	5	6	7	8	9	10	Aver. for row
1	7857	207	206	209	210	203	209	209	210	210	210	208
2	7853	202	207	209	208	208	209	210	208	210	210	208
3	7902	209	199	199	197	199	199	200	200	200	210	201
4	7695	210	199	189	176	190	183	198	200	199	210	195
5	7248	197	186	187	153	184	186	181	187	187	197	184
6	7448	197	197	196	196	195	196	197	197	198	198	197
7	7346	195	197	181	193	188	197	197	197	197	197	194
8	7260	185	197	183	197	196	167	197	198	197	197	191
9	7030	177	193	168	192	191	189	192	171	192	192	186
10	7283	192	176	192	192	192	192	192	192	192	192	190

Average input for all burners 26,120 BTU/hr
Total input in whole field 19,530 10^6 BTU

Total number of hours the burners were off during test

Row	Burner no. 1	2	3	4	5	6	7	8	9	10	Σ (row)	% of tot. time
1	269	275	160	138	385	166	163	147	148	147	1998	2.48
2	395	246	179	192	200	169	153	205	149	147	2035	2.52
3	158	153	168	233	147	167	130	134	128	128	1546	1.92
4	137	139	507	965	587	706	167	124	164	124	3620	4.50
5	84	100	264	1838	170	91	264	76	64	82	3033	4.01
6	72	100	142	122	171	113	80	104	65	63	1032	1.37
7	111	146	729	234	385	91	113	86	77	77	2049	2.71
8	500	82	540	66	145	1270	104	66	67	74	2914	3.86
9	653	88	1031	105	135	194	60	911	58	58	3293	4.47
10	64	665	64	64	60	62	82	64	67	69	1261	1.71
											22.781	2.95

casing, before they flowed upwards through the annulus space between the supply tube and the casing, to the surface. The supply tube was designed so that the gas velocity would be high enough to prevent "flashback" into the fuel gas supply.

This and other burner types are described on Figs. L2-100, 101 and 102.

The next burner tested, Type B, had a small jet, inserted between the supply tube and the cone, designed to recirculate a certain amount of exhaust gas for better heat distribution along the burner. The jet could be adjusted to provide the desired exhaust gas recirculation rate.

The Type C-burner had a thermal shield ("hood"), consisting of a concentric steel tube, placed around the cone and the upper part of the burner tube, with the purpose of shielding the burner well casing from part of the heat, radiated from the hottest part of the burner. The Type D-burner had two such concentric shields of different lengths.

The Type E-burner had a thermal shield called a "combined hood and burner tube", considerably longer than the burner tube.

All burners with hoods were built with jets for exhaust gas recirculation.

In order to extend the length of the heated interval in some tests, a Type B-burner was moved up and down in the well at regular time intervals.

Later a new burner type was developed, consisting of a type-A burner with a fluidized bed of sand in the annulus between the burner tube and the casing. The fluid sand bed was intended to act as a heat transfer medium, distributing the heat uniformly along the casing. Sands of different origins and compositions, with grain sizes ranging from 6 to 100 mesh were tested. In the test descriptions the amounts of sand used are given as the heigths of the sand bed, when resting on the bottom of the empty burner casing.

Fuel and air supply

Commercial propane was used as fuel in all tests, except in one test, where a burner was run with produced gas from a seven-burner test. It was found that a burner with this fuel was easier to start and showed a good flame stability within wider ranges of heat input than propane-burners did.

Air was supplied from piston-type compressor, or (in the hundred-hole test) from a positive displacement blower.

During most of the single-burner and seven-burner tests (all except L72, 8, 8A, 105-119) air and propane were mixed in a Lindell type mixing valve, with proportioning gates for both gases.

The amounts of propane and air to tests L105 through L119 were controlled individually with needle valves.

In order to maintain constant air-fuel-ratios, the propane pressure was kept the same as the air pressure by a propane pressure regulator, which was controlled by the air pressure. No corrections were made for variations in air and propane temperatures (e.g. between day and night).

The air and propane flows were measured with rotameters immediately before the mixing valve or the needle valves.

For the hundred-hole tests (L8 and L8A) and the concurrent seven-hole test (L72) air and propane were controlled and mixed in any desired proportions by a Honeywell-Brown ratiocontroller. The gas flows were measured with orifices.

In all tests a stoichiometric ratio between air and propane (24 to 1) was maintained as closely as possible. As a check Orsat analyses of the exhaust gas were made from time to time. The deviations from ideal conditions calculated from the O_2-content of the exhaust gas, were only occasionally more than 2 %.

Heat inputs

The amount of fuel, supplied to any burner during a test or a certain part of a test was kept as constant as possible. Different tests were run with heat inputs, varying from 15,000 to 35,000 BTU/burner-hour. As it was found that the optimum input to a certain burner was related to the length of the burner tube, also the input divided by this length (BTU/hr, ft burner tube) is given in most of the test tables in this report. The accuracy of the given input figures is estimated to be within $\pm$ 5 %.

It should be noted that all heat input figures are calculated from the gross heat value of the supplied propane (2509 BTU/ st cuft). No correction has been made for the heat content of the outgoing exhaust gas. The temperature of the exhaust gases when leaving the casing was measured only in a few cases, but was probably between 150 and 300°F in most of the tests. In addition to that, heat is also lost via the exhaust gas to the overburden.

Temperature measurements

The temperatures in the formation were measured with thermometers, mounted in holders, made of 1-ft long, concentric pieces of 1/4-inch and 1½-inch steel pipe. The holder, being attached to a thin steel wire, running over a calibrated depthmeasuring wheel to a reel, could be lowered to any desired depth in the formation, inside the casing of the temperature well. In order to attain temperature equilibrium with the surrounding formation, the holder with the thermometer was left at the desired level for at least 2 hours. The high thermal lag of the thermometer holder ensured accurate readings after the holder had been brought up to the surface.

Temperature wells were located at different distances from the burner wells.

Construction materials

The high temperatures encountered in some parts of the underground equipment, made it necessary to use heat-resistant construction material. In the single-burner tests, the emphasis was put on the different parts of the burner. The following materials were tested:

In the supply pipe:	carbon steel and 18/8 stainless steel,
" " cone:	25/20 stainless steel (plate);
	25/12 stainless steel (cast);
	25/0 stainless steel, "Fernox", (cast);
	and an aluminum-iron-alloy, "Kanthal", (cast);
" " burner tube:	18/8 stainless steel, 25/20 stainless steel,
	25/12 stainless steel and carbon steel
	(lower end of burner tube only).

It was found necessary to test not only carbon steel casings, but also the following steel alloys:

25/12 stainless steel (cast, "Thermalloy")
9 % Cr, 1 % Mo, 0.75 % Si.
2.5 % Cr, 1 % Mo.
5 % Cr, 0.5 % Mo, 1.5 % Si.

RESULTS OF THE TESTS

Burner efficiency

A suitable burner should supply the same amount of heat to the formation from each unit of its entire length, thereby establishing a uniform temperature along the burner casing, assuming that the tar sand layer is homogenous. An uneven distribution of heat, resulting in local temperature peaks at the casing is undesirable. In every test, where the casing failed, this was due to an uneven heat distribution.

In order to rank the different burners tested, the following "characterization numbers" were used where the term "temperature" denotes the increase above the ambient temperature:

1. $T_{Avg.}/T_{Max}$. The average temperature along the burner tube divided by the maximum temperature is called the burner efficiency. The average temperature is a measure of the total heat transferred to the casing and if it is equal to the maximum temperature, then the rate of heat transfer is uniform along the entire burner tube and the efficiency is 100 %.

2. L_{80}. This is the length of the interval which is heated to, or above, a temperature equal to 80 % of the maximum temperature.

3. L_{50}. This is the length of the interval heat to at least 50 % of the maximum temperature.

4. L_{80}-L_{50}. The difference in these two lengths is a measure of the amount of heat being transferred outside the desired interval. If L_{80}-L_{50} is zero the temperature curve would have a rectangular shape and very little or no heat would be transferred above the desired interval.

Results

For the results and conclusions, obtained in the individual tests, reference is made to the detailed descriptions in the Appendix. The chronological order of the tests is shown on Fig. LO-800.

Group 1. Short-time tests with single burners without sand

The burners of each kind, showing the highest efficiencies, were:

Test No.	Burner Type	Burner tube length, ft (fr. cone to bottom)	Heat input BTU/hr	Exhaust gas recirc. %	Effic. $\frac{T_{Avg.}}{T_{Max}}$ %	L_{80} ft
L22-1	A	23½	22,500	0	32	3
L22-2	B (jet)	23½	22,500	15	32	3
L22-7	C(jet+1 hood)	23½	22,500	15	37	3
L22-8	D(jet+2 hoods)	23½	22,500	15	41	4
L22-17	E(jet+long hood)	21	20,000	25	47	4
L22-18	E -"-	21	35,000	15	49	4
L22-19	E -"-	14½	20,000	25	58	3½

Thus with exhaust gas recirculation and thermal shields ("hoods") a slight improvement in heat distribution was obtained. The better efficiency of the last-mentioned E-burner over the other E-burners was due to its shorter length and did not signify an improvement from the point of view of heat distribution over a longer distance.

Group II. Long-time tests with single burners without sand

Test No.	Burner Type	Burner tube		Heat input BTU/hr	Heat distribution data			
		Diam. inch	Length ft		T_{Max} °F	$\frac{T_{Avg}}{T_{Max}}$ %	L_{80} ft	$L_{80}-L_{50}$ ft
L2	A	3/4	26.5	30,000	425	29	4.5	3.5
L3	"	"	27	"	565	39	6.5	4.5
L6	"	"	17	25,000	490	35	3.5	3
L4	"	1	27.5	40,000	480	31	4	4.5
L5	"	"	"	"	460	28	4	4
L6	B	3/4	17	25,000	480	33	3.5	2
L4	"	1	27.5	30,000	360	50	8	11
L101	"	"	20	"	290	62	5	23
L5	C	"	27.5	"	400	40	7	9.5
L4A	E	"	"	34,000	505	38	5.5	6
L61	"	"	21	20,000	680	29	3.5	3.5

In contrast to the short-time tests the above tests showed that the type B and C burners gave the highest efficiency and the longest L_{80} heated intervals while the typer E burner did not show any improvement over the type A burner.

Group III. Seven-burner tests with burners without sand

Burners with hoods were placed in the center and in the six corners of a hexagonal pattern with 4-ft sides. During the first 29 days of operation, 3/4-inch diameter burners were used with up to 15 % exhaust gas recirculation. These burners were found to be less stable in operation than the 1-inch burners used during the next 78 days. However, repeated failures in burner cones and casings occurred due to the high local temperatures that were reached towards the end of the test. In multiburner tests such as these, the casing and formation temperatures are higher than they would be in single burner tests under the same conditions. The higher temperatures are caused by interference between wells, i.e. several burners are heating an area which would otherwise be heated by only one.

In another seven-hole test with the same well pattern, 1-inch burners with exhaust gas recirculation but without hoods were moved up and down. Also here the casings started to burn off after about 60 days' heating, while the burners operated without failures up to about 140 days. Although the burners were moved over an interval of 30 feet, temperature measurements in the formation, shortly before the test was finished, showed a good heat distribution over an interval of only about 20 feet.

Group IV. 100-burner and 48-burner tests with and without sand

A larger-scale test, consisting of 10 rows of burners with 10 burners in each row, was run between July 1956 and May 1957. The burners were arranged in a triangular pattern with 8 foot spacing. The heat input varied between 17,000 and 20,000 BTU/h-hr and the burners were moved up and down in the casing every 4 hours. Recirculation of exhaust gas was used during the first part of the test only. Most of the casings failed in spite of the fact that the burners were moved up and down. The test was continued on a limited scale during the next 225 days. Burner casings of different materials were tested in 48 of the wells with moving burners without sand as well as sandburners.

The results showed that the requirements on heat-resistant materials for burner casings were less severe with sand burners than with burners without sand. It appeared possible that plain carbon steel casings could be used with sand burners.

The temperature curves during the last part of the test, when sand burners were used, showed a great improvement in heat distribution over those of the first test period.

Group V. Single-burner tests with sand burners

A number of tests with 5 to 35 ft long sandburners were run in order to establish an approximate basis for a further systematic study of this burner type.

A summary of the results is given in Table LO-700.

The tests showed that the following factors affect the heat distribution:

1. Sand size.

2. Heat input.

3. Amount of sand.

4. Length of burner.

Heated intervals L_{80} in the range of 30 to 34 feet were obtained with a 20 ft burner tube at heat inputs from 22,000 to 30,000 BTU/hr and with a 25 ft burner tube at heat inputs from 25,000 to 32,000 BTU/hr. Sand levels were in the range of 6 to 10 feet.

The temperature curves on Fig. LO-401 illustrate the improvement of the heated interval L_{80} from a type A burner to a sand burner.

Condensed results of Sandburner tests

Test No.	Burner length ft	Extension tube ft	Heat input 10^3 BTU/hr	Sand: ft of casing	Sand: Size mesh	T_{Avg}/T_{Max} %	L_{80} ft	$L_{80}-L_{50}$ ft
Sand size range: 60 - 100 mesh								
108B	5	-	20	1.2	60-100	98	7	3
C	"	15	"	"	"	95	7	3
D	"	"	"	2	"	98	7	4.5
106B	10	-	"	"	"	87	14	3.5
107C	15	-	30	"	"	77	6.5	14
E	"	-	"	1.5-3	"	74	6.5	15.5
D	"	-	20	"	"	57	3	6.5
B	"	-	"	2	"	37	2	2
Sand size range: 20 - 60 mesh								
113B	10	-	25	2.5-5	40-60	92	11	5
A	"	-	30	5	"	94	11	6
115F	15	-	25	7-10	20-40	89	23	8.5
C	"	-	"	7	40-60	85	17	3
107G	"	-	30	2.2-3	"	90	16.5	7
115A	"	-	20	5	"	81	13	2.5
E	"	-	"	10	20-40	82	13	8
110A	"	25	"	5	40-60	90	16	4
118A	20	-	22	8-10	20-40	94	33.5	4
116A	25	-	25	5.5-10	40-60	86	28	5.5
111F	"	15	32	"	20-40	90	34	6
B	"	"	27	6-8	40-60	90	31.5	8
A	"	"	21	10	"	78	15	11
112B	35	5	25	8-10	"	73	16	15

cont.

cont.

Test No.	Burner length ft	Exten-sion tube ft	Heat input 10^3 BTU/hr	Sand: ft of casing	Sand: Size mesh	T_{Avg}/T_{Max} %	L_{80} ft	$L_{80}-L_{50}$ ft
Sand size range: 8 - 30 mesh								
115K	15	-	30	7-10	12-14	92	20.5	18
J	"	-	25	9-10	10-12	92	20	17.5
H	"	-	20	8.5-10	10-12	90	18	3
107F	"	-	30	1.4-2	16	83	11.5	11.5
118D	20	-	30	6.5-10	10-12	91	33.5	5
C	"	-	25	8.2-10	"	92	32	5.5
F	"	-	30	7.5-10	8-12	96	31	4
117A	"	-	30	9-10	"	71	29	5
118B	"	-	20	"	10-12	87	24.5	8.5
116F	25	-	25	8-10	12-14	93	32	7
G	"	-	30	9-10	"	91	29	9
119A	"	-	"	"	"	86	28	12.5
B	"	-	"	"	8	84	26.5	14.5
116D	"	-	"	7-10	14-16	79	20	11
E	"	-	20	9-10	10-12	78	19	15
111C	"	15	27	7-9	10-30	79	24	6
D	"	"	32	7-8	"	86	24	10.5
E	"	"	38	6-7	"	85	24	12

Oil and gas production

The permeability of the tar sand being low, the flow of products towards separate gas wells was restricted in tests where these wells were used. Pressures of up to 6 psig built up around the burner wells without any vapors reaching the gas wells 4 ft distant. The permeability changes, however, with temperature and in the multiburner tests, after higher average formation temperatures had been reached, a flow of vapors probably took place between different parts of the test formation.

The different types of gas well completions did not show any significant differences in performance even though the "gravel-packed" gas wells in the first seven hole tests showed a slightly smaller oil production and higher gas production than the "open" wells did. In the 100- and 48-burner tests where a total of 371 bbls of oil was produced, open, concentric gas wells were used.

Some difficulties were met in avoiding plugging, in the gas wells and product lines by tar, produced during the first part of each test. Also the contamination of the produced oil with tar resulted in water - oil emulsions, which were hard to break. For the bigger test units, an emulsion treater, working at 150°F and with addition of de-emulsifying agents, was used with satisfactory results.

Some samples of the produced oil were analyzed. One precision fractionation was made. The oil was aromatic, unstable and contained 2½ %, by weight, of sulfur. Complete analyses are shown in the Appendix.

Analyses of samples of the produced gas showed:

H_2S	6 - 12 % by volume
CO_2	3 - 15
H_2	24 - 46
Olefins	4 - 18
Paraffins	24 - 45

Net heat value 800 - 1000 BTU/st cuft.

Complete analyses are found in the Appendix.

Acknowledgement

The work reported was carried out by a team of Mr. William J. Shirley, Mr. Malte O. Eurenius and the author. Grateful acknowledgement is due Dr. Robert E. Helander for valuable and skillful assistance in the preparation of this report.

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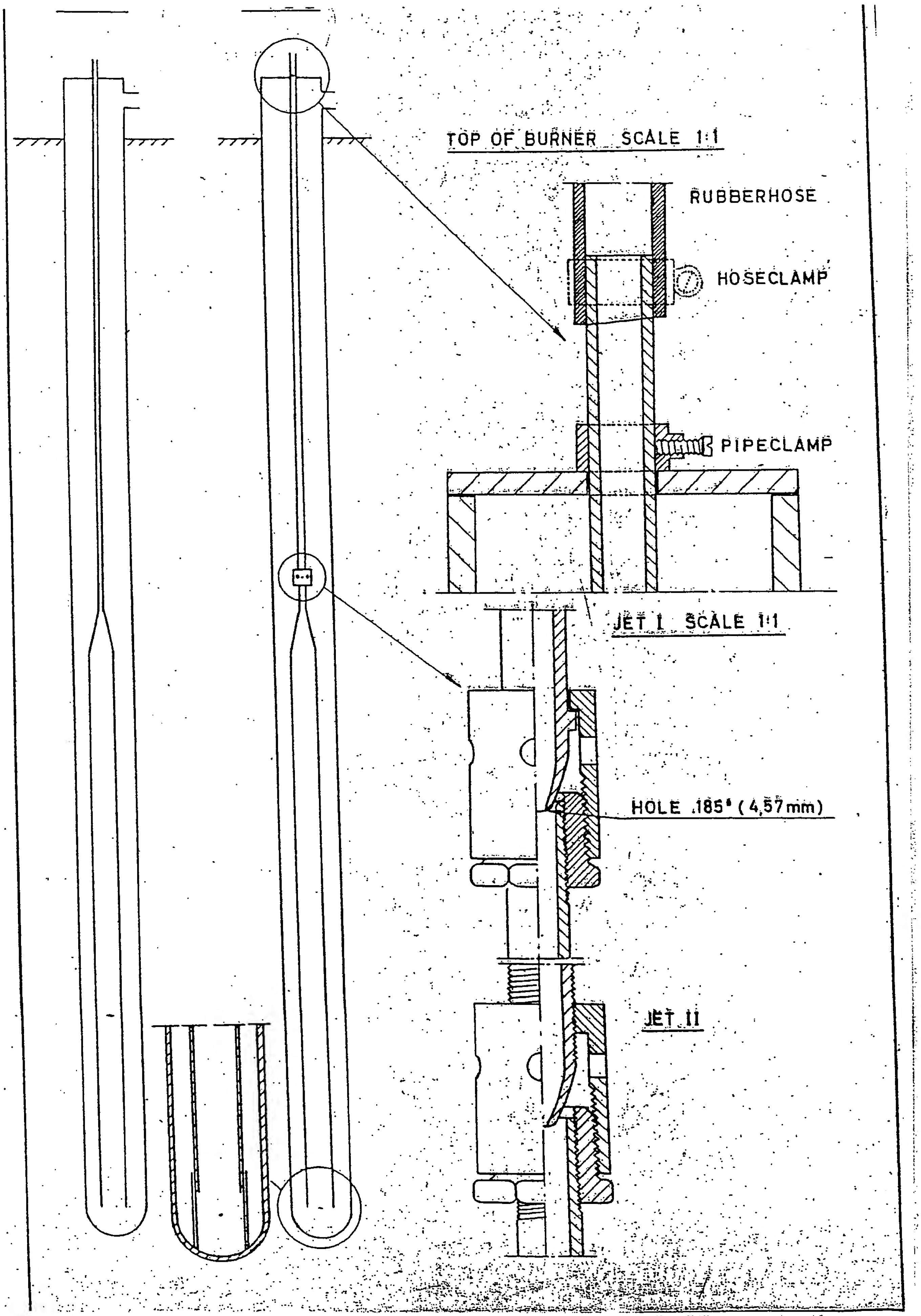
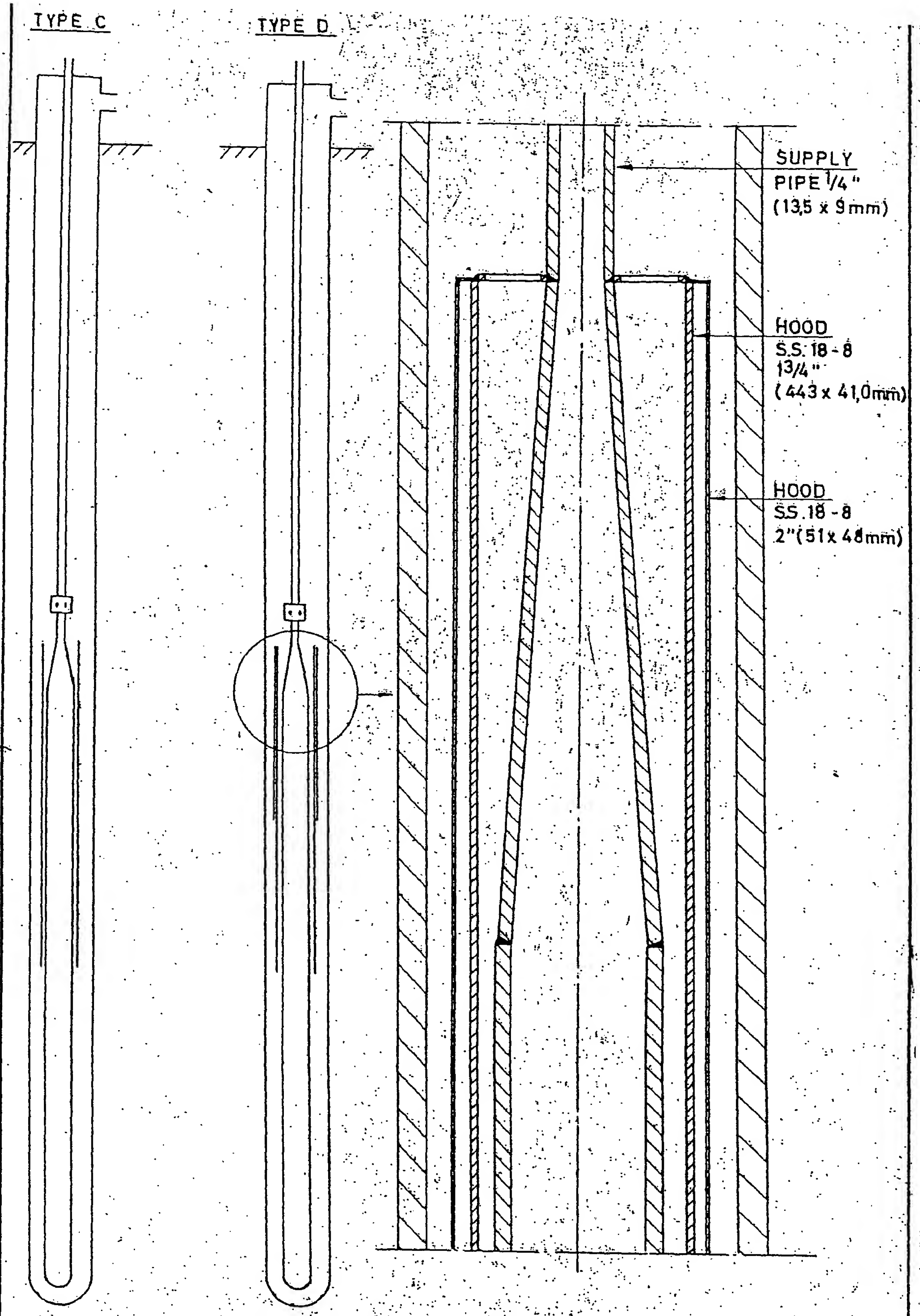
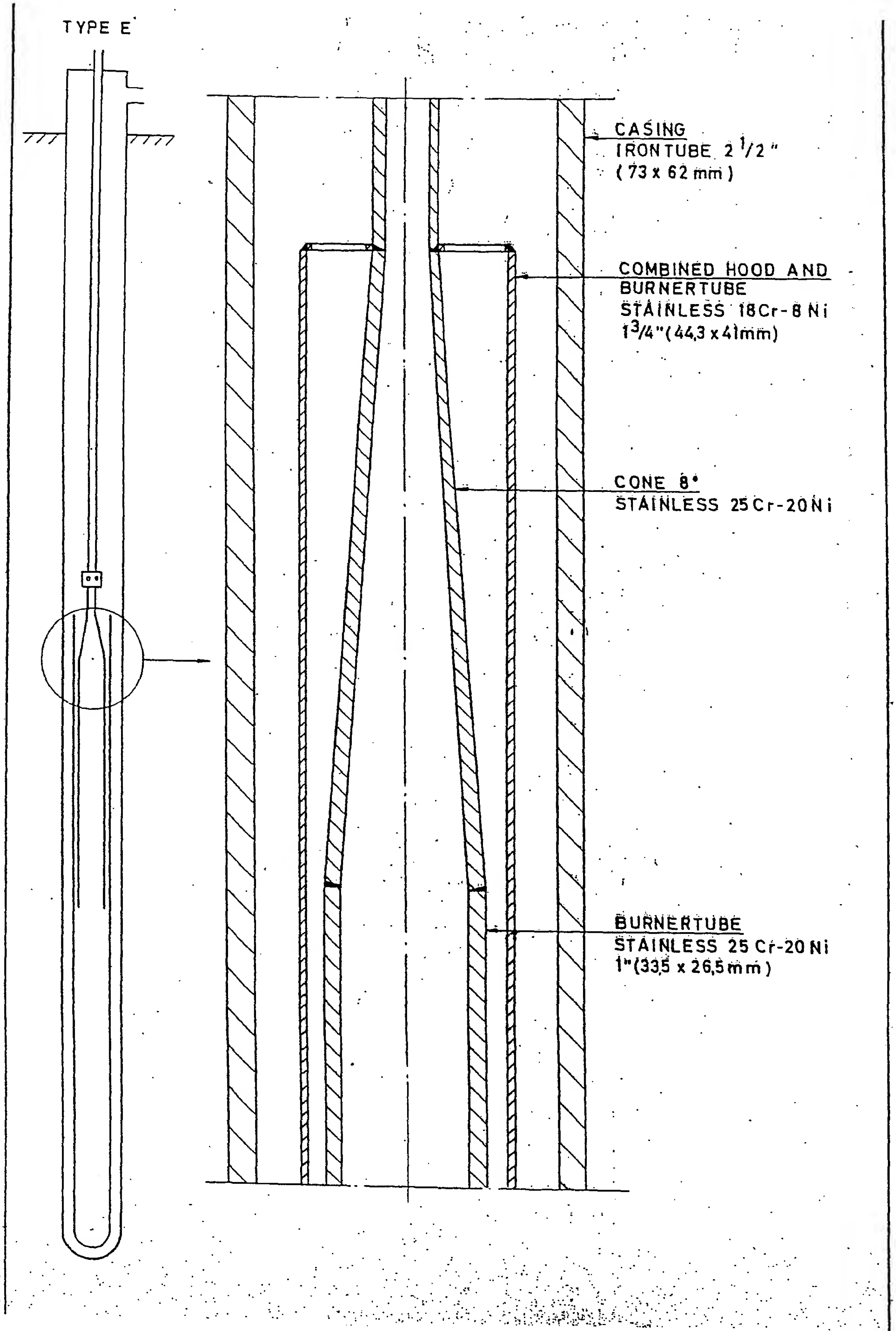


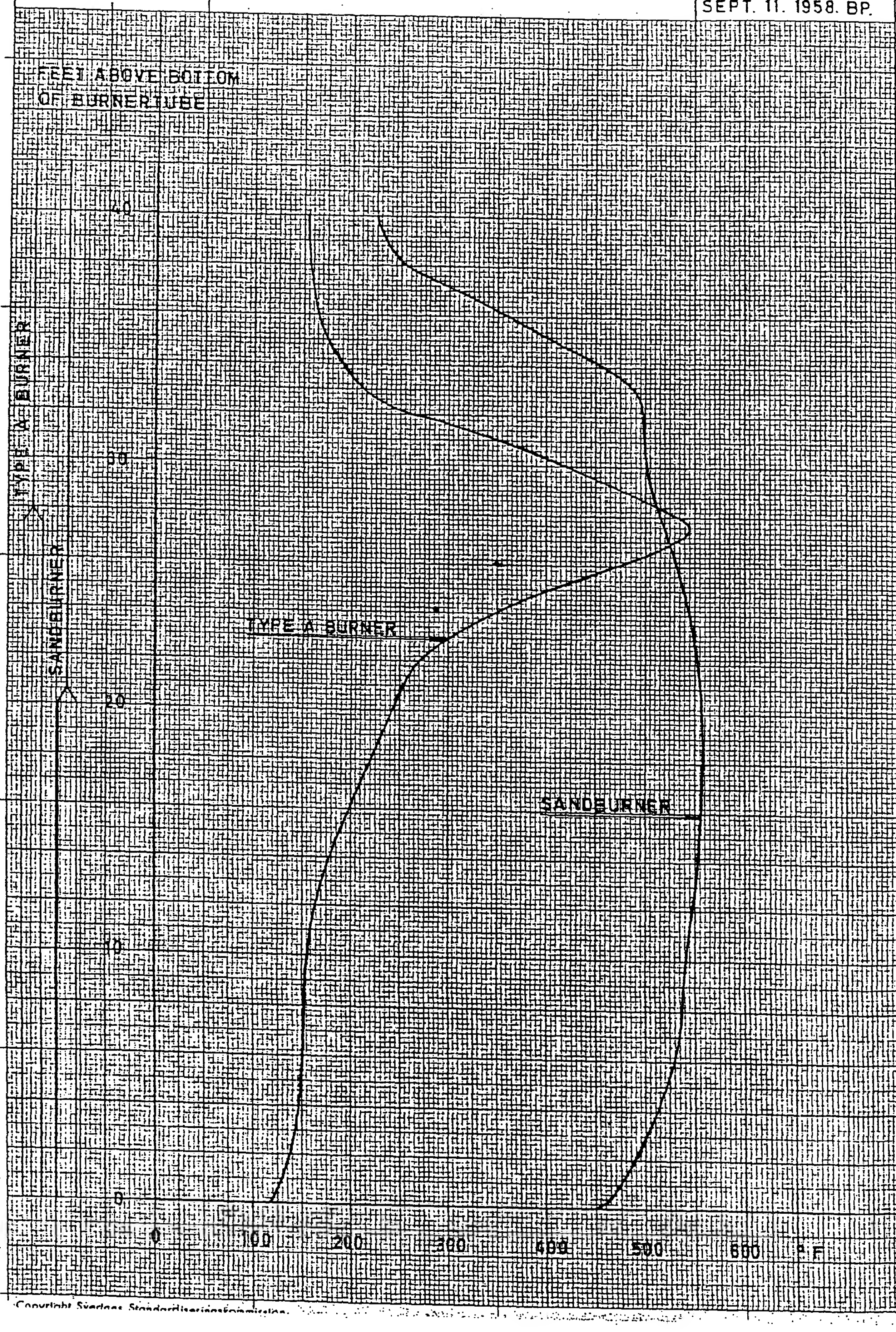
TOP OF BURNER SCALE 1:1
RUBBERHOSE
HOSECLAMP
PIPECLAMP
JET I SCALE 1:1
HOLE .185" (4,57mm)
JET II



TYPE C
TYPE D
SUPPLY
PIPE 1/4"
(13,5 x 9 mm)
HOOD
S.S. 18 - 8
1 3/4"
(44,3 x 41,0 mm)
HOOD
S.S. 18 - 8
2" (51 x 48 mm)


TYPE E
CASING
IRONTUBE 2 1/2"
(73 x 62 mm)
COMBINED HOOD AND
BURNERTUBE
STAINLESS 18Cr-8Ni
1 3/4" (44,3 x 41mm)
CONE 8°
STAINLESS 25Cr-20Ni
BURNERTUBE
STAINLESS 25Cr-20Ni
1" (33,5 x 26,5mm)

LO - 401.
SEPT. 11. 1958. BP.

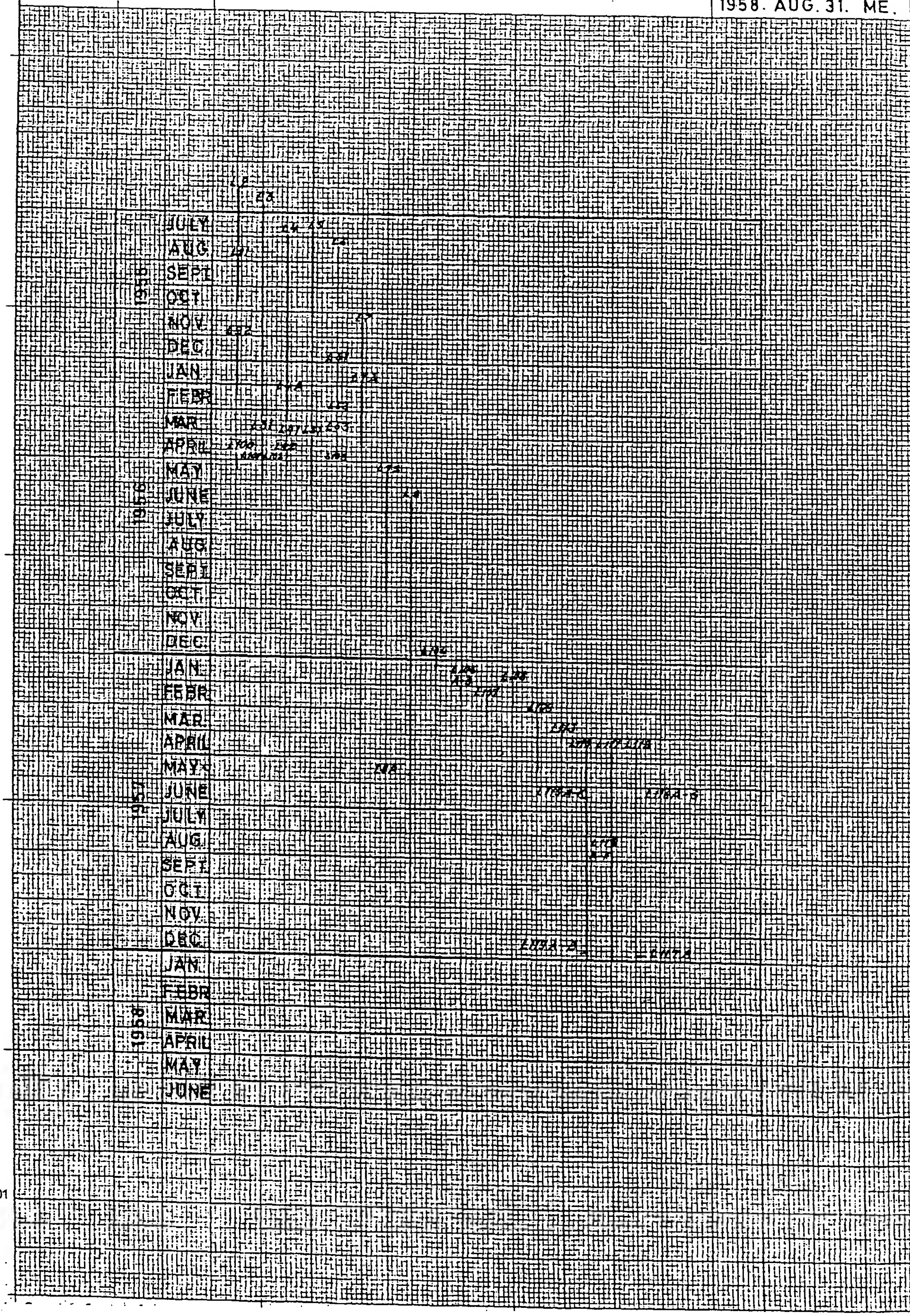


523 A4
73 25 01
1 x 1 mm
ESSELTE
4446

IN SANTA CRUZ CALIFORNIA

LO - 800

1958. AUG. 31. ME.



HUSKY OIL COMPANY

Cody, Wyoming

TECHNICAL SERVICE DEPT. REPORT | No.: 56-H-25

Subject: SANTA CRUZ DISTILLATE | Date: May 21, 1956

To: M. Westfall | From: J. R. Hartwig

SYNOPSIS

Analysis of the various cuts obtained from a precise fractionation of a sample of Santa Cruz distillate indicates that this material is an unstable mixture of complex olefins, cycloparaffins, cyclo-olefins, terpenes, sesquiterpenes and high boiling aromatics. The analysis does not indicate the presence of significant amounts of commercially valuable chemicals such as benzene or toluene. This material would not be a suitable charge stock for a chemical plant for the recovery of such chemicals. Catalytic desulfurization and reforming of this stock would probably be beneficial. However, the extent of this improvement is not known and further laboratory work along these lines is recommended only if the process seems commercially feasible.

INTRODUCTION

On March 15, 1956, M. R. Westfall requested that the Technical Service Department run a precision fractionation on samples of Santa Cruz distillate. The first sample of this distillate was received on March 20 and a second on April 2. Our usual Hempel analysis of these samples were reported in reports 56-S-4 and 56-S-8. The precise fractionation of the Santa Cruz distillates could not be performed until now because we had to order the special equipment required for this type of distillation.

PURPOSE

The purpose of this work is to determine if Santa Cruz distillate would be a suitable feed stock for a chemical plant producing aromatics such as benzene, xylene, toluene, napthalene, or any other chemicals of commercial value.

METHOD

The gasoline, naphtha, K.D. and L.G.O. cuts of the atmospheric Hempel distillation of Santa Cruz distillate number LA7 were combined for the charge for a precise distillation. This distillation was performed in a precise fractionation assembly purchased from the Todd Scientific Co. A 25 mm. I.D. column with a 90 cm. packed length was used. This column has 42 theoretical plate at total reflux. A reflux ratio of 25 to 1 was used. Cuts were taken at one, five, or ten °C. increments throughout the distillation. These cuts were analyzed by refractive index, specific gravity and also by means of published correlations relating to molecular weight and ring content of hydrocarbon mixtures.

RESULTS

In TABLE I the results of the precision fractionation are shown. These data are also shown as curves on Figures 1, 2 and 3.

cont.

Cut No.	Cut %	Total %	Boiling Range @ 760mm°C	Boiling Rang @ 760°F./mm	Specific Gravity@ 20°C	R.I. @ 20°C, No.	Molecular Wt. (Avg.)	C.I.	Rings Per Molecule
[illegible]	.9	1.86	60-65	140-149	—	1.3950	82	---	.3 - .5
[illegible]	.78	2.64	65-70	149-158	.702	1.4150	85	20	.4 - .8
[illegible]	.67	3.31	70-75	158-167	.748	1.4260	85	41	.5 -1.0
4	.70	4.01	75-80	167-176	.767	1.4308	88	46	.5- 1.1
5	.60	4.61	80-85	176-185	.738	1.4233	90	27	.45-.95
6	.70	5.31	85-90	185-194	.709	1.4169	93	15	.38- .8
7	1.02	6.33	90-95	194-203	.704	1.4156	95	11	.38- .8
8	1.06	7.39	95-101	203-214	.742	1.4244	99	28	.45- .9
9	1.30	8.69	101-106	214-223	.774	1.4328	100	41	.5 -1.1
10	2.91	11.60	106-116	223-241	.758	1.4283	102	30	.45- .9
11	2.14	13.74	116-126	241-256	.732	1.4220	110	15	.55-1.2
[illegible]	4.40	18.14	126-136	256-277	.782	1.4353	113	36	.45-1.0
[illegible]	4.16	22.30	136-146	277-295	.790	1.4388	120	37	.5 -1.0
14	2.79	25.09	146-156	295-313	.794	1.4398	121	36	.5 -1.0
15	5.30	30.39	156-167	313-333	.809	1.4472	130	40	.55-1.2
16	2.47	32.86	167-172	333-341	.819	1.4513	132	43	.55-1.3
[illegible]	1.38	34.24	172	341	.823	1.4532	—	45	—
[illegible]	1.50	35.74	172-177	341-350	.822	1.4530	135	43	.57-1.3
19	3.48	39.22	177-187	350-368	.828	1.4558	140	44	.6 -1.4
20	3.37	42.59	187-197	368-386	.836	1.4594	144	46	.6 -1.4
21	4.55	47.14	197-207	386-404	.849	1.4680	150	50	.75-1.6
22	2.90	50.04	207-217	404-422	.856	1.4716	151	50	.8 -1.7
23	4.70	54.74	217-222	422-431	.849	1.4673	161	46	.7 -1.6
24	3.35	58.09	222-223	431-433	.849	1.4680	160	45	.75-1.6
25	5.15	63.24	223-230	433-446	.853	1.4690	163	45	.7 -1.6
26	3.00	66.24	230-237	446-458	.850	1.4715	171	44	.7 -1.7
[illegible]	4.80	71.04	237-243	458-469	.861	1.4736	178	48	.85-1.8
[illegible]	4.40	75.44	243-253	469-487	.868	1.4770	181	50	.85-1.9
[illegible]	5.65	81.09	253-258	487-496	.872	1.4801	186	50	.9 -2.0
30	4.50	85.59	258-268	496-514	.879	1.4842	191	52	.95-2.2
31	1.44	87.03	268-270	514-518	.885	1.4878	194	54	1.0-2.3
32	3.67	90.70	270-278	518-532	.891	1.4908	199	55	1.1-2.4
33	2.69	93.39	278-288	532-550	.894	1.4933	206	55	1.1-2.5
[illegible]	1.99	95.38	288-295	550-563	.897	1.4944	212	56	1.1-2.6

Residue 3.0%

Loss 1.62%

* Startup sample after overnight shutdown.

The cuts used for the above distillation amount to 74.4 percent of the total Santa Cruz distillate sample number LA7. These cuts were obtained from the distillation reported in 56-S-8.

cont

DISCUSSION

The first six cuts of the distillation were water-white, the remaining cuts were all ıted a reddish color. After standing overnight all cuts turned a deep red to purple ɔr and cuts 10 through 34 had a gum form on the bottom of the sample bottles. This evidence, aswell as the analysis of the cuts presented in TABLE I and Figures 1, 2, and 3, indicates the unstable and unsaturate nature of the Santa Cruz distillate. The data on Figure 3 indicate that this distillate is composed primarily of normal and iso-olefins, cyclohexanes, cyclohexenes and terpenes. Also, in view of the method used to recover this distillate at Santa Cruz, a considerable quantity of oxygen and nitrogen hydrocarbon derivatives may be expected. These compounds also contribute to the discoloration and unstableness of this distillate.

In Figure 1, the plot of boiling point versus percent distilled indicates the complexity of this material and the absence of any large cut boiling is a narrow range. There is a slight indication of plateaus at 220°C, 255°C, and 270°C boiling points. However, I know of no hydrocarbons boiling in these ranges that have commercial value. Furthermore, the percent of the total crude in these ranges is too small for commercial production. Also ıown on Figure 1 is a plot of refractive index versus percent distilled. The refractive indices shown are for each cut boiling within the ranges indicated.

On Figure 2 a plot of refractive index versus boiling point is shown. Literature data for various types of pure hydrocarbons are also shown on this curve. This curve again indicates that the Santa Cruz distillate is a complex mixture. Small amounts of benzene and toluene may be present. For instance, the R.I. - Boiling Point plateaus tend toward a peak in the boiling range for benzene and also for toluene. However, these peaks are slight and the percentage of pure benzene and toluene would be very low.

The stability of this distillate would no doubt be improved by catalytic hydrogenation desulfurization treatment. This would remove the sulfur, nitrogen, and oxygen compounds hydrogenate the unsaturates. After this treatment catalytic reforming could be used to increase the aromatic content. However, these treatments would be costly and it is doubtful that the beneficial affects of these treatments would produce a charge stock for a chemical plant which would yield enough products such as benzene or toluene to be profitable.

.NCLUSIONS

1. The Santa Cruz distillate is a complex mixture of olefins, cycloparaffins, cyclo-olefins, terpenes, and high boiling aromatic compounds.

2. The distillate is very unstable due to the unsaturation and the presence of nitrogen, oxygen, and sulfur compounds.

3. The distillate does not contain appreciable amounts of any one compound and attempts to extract a particular compound would give very low yields.

4. This analysis indicates that the distillate is unsuitable as a feed stock for a chemical plant. Catalytic desulfurization and reforming would beneficiate this distillate but the economics of this treatment should be studied carefully before considering a commercial venture.

RECOMMENDATION

Further laboratory work on Santa Cruz distillate may be desirable to investigate the 'ect of desulfurization on this stock. Therefore, I recommend that we run a sample of

cont.

this distillate through our laboratory pilot plant desulfurizer at the first convenient opportunity if the process for extracting this material from the tar sand at Santa Cruz proves economically feasible.

APPROVED F. B.
F. B. ODASZ

cc: Brummond
Vokac
VII-17
Hartwig (2)

cont.

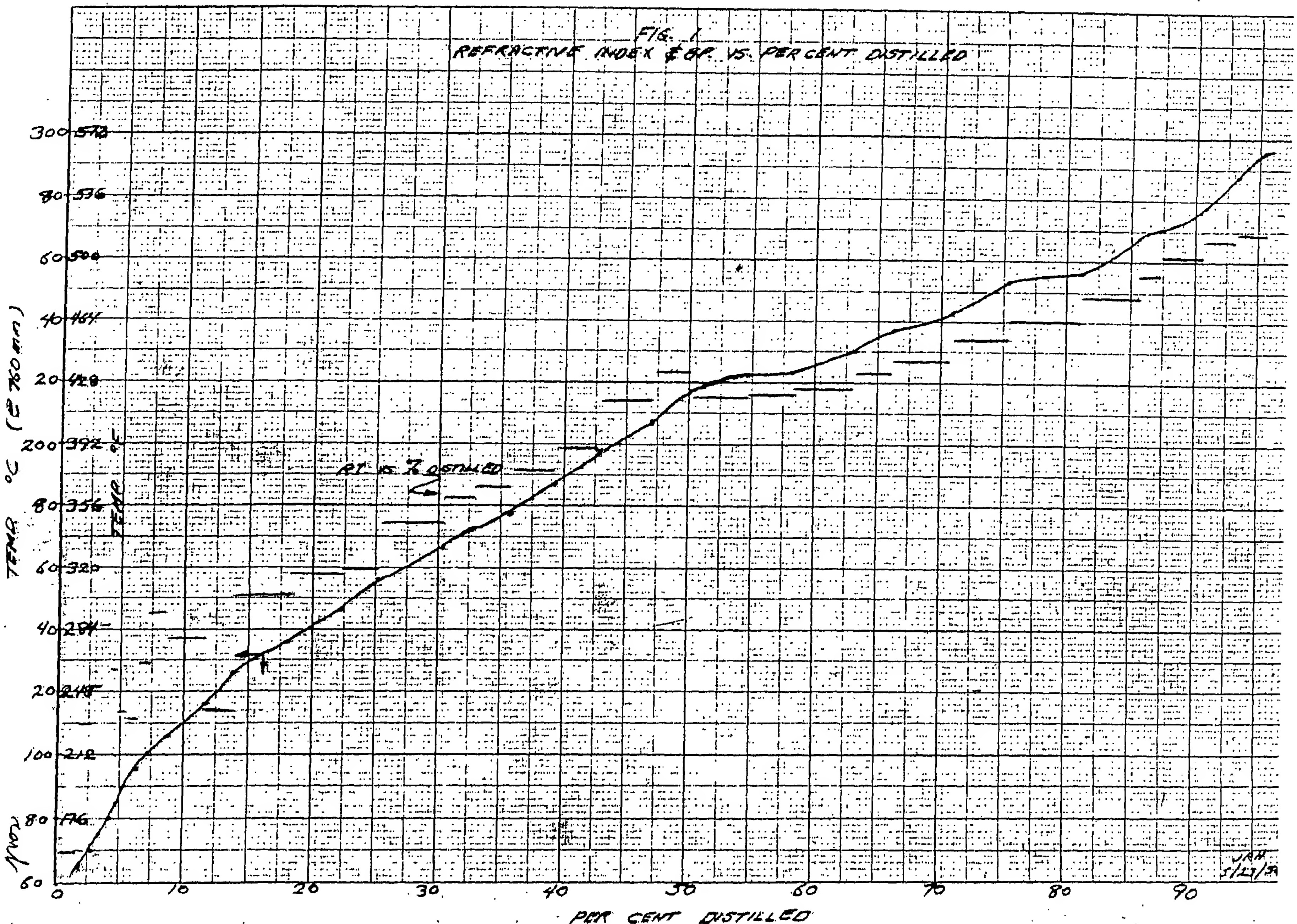
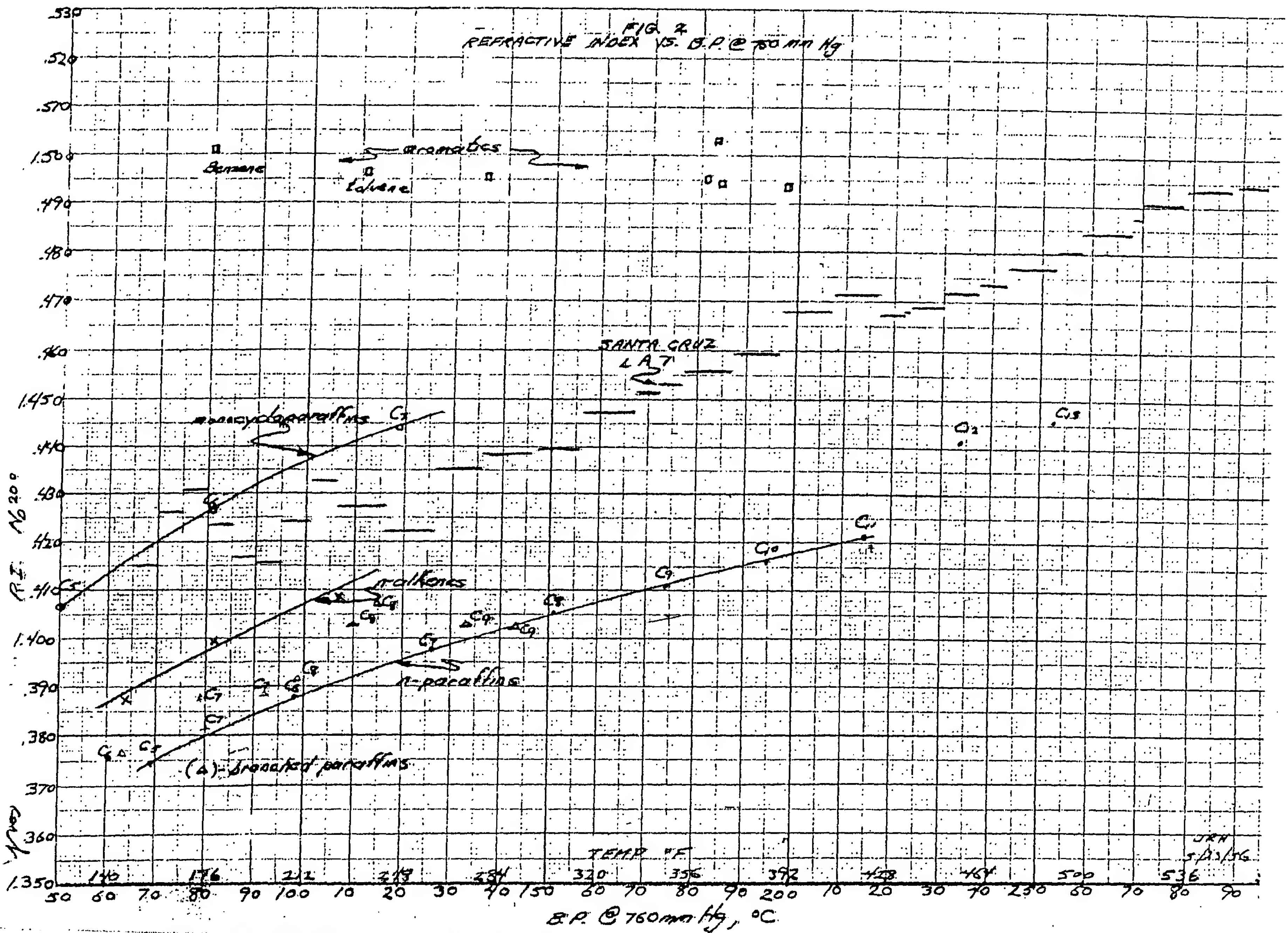
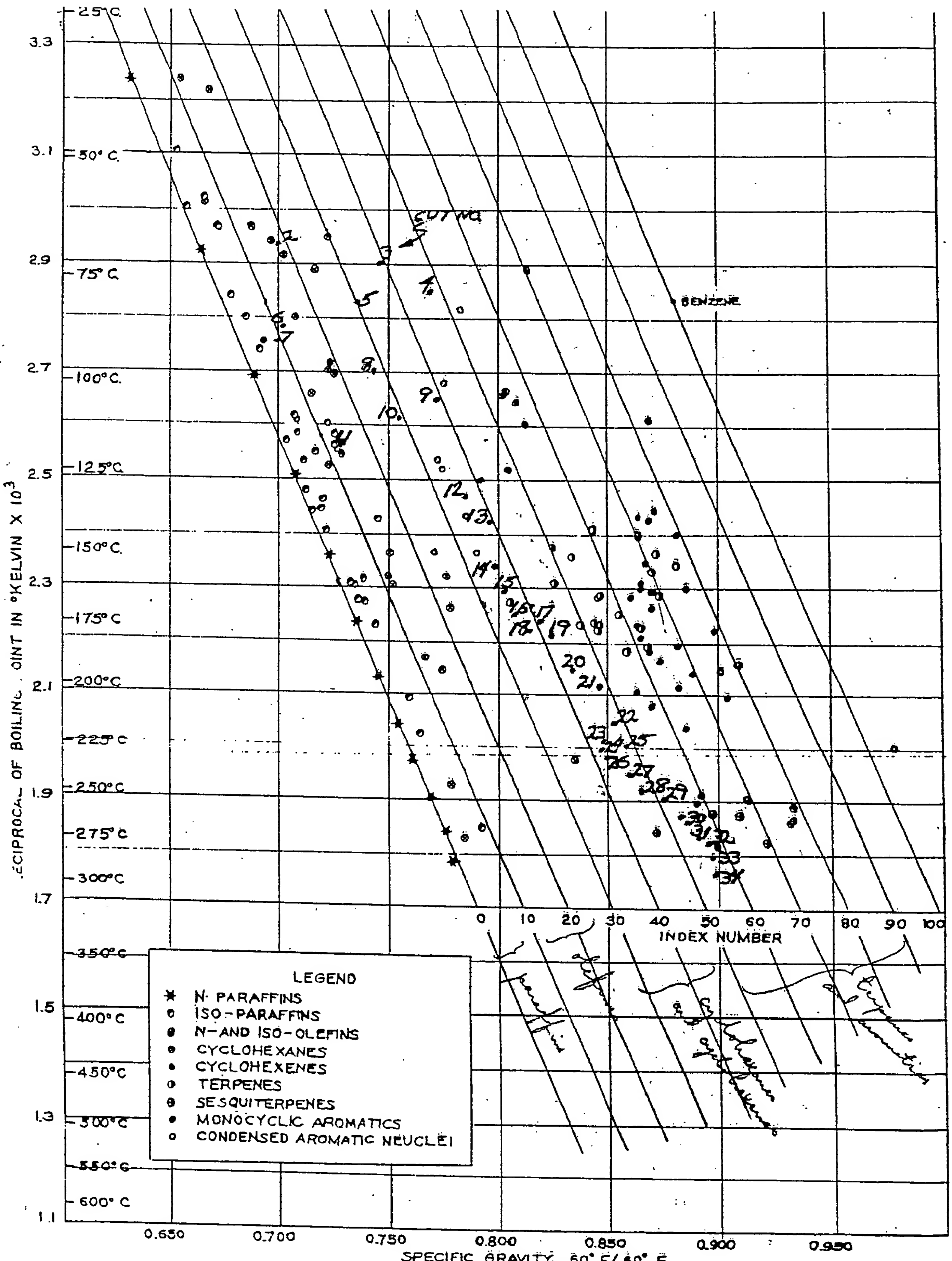


FIG. 1
REFRACTIVE INDEX & BP VS. PERCENT DISTILLED
RI VS % DISTILLED
TEMP °C (@ 760 mm)
TEMP °F
300 572
80 536
60 500
40 464
20 428
200 392
80 356
60 320
40 284
20 248
100 212
80 176
60
0
10
20
30
40
50
60
70
80
90
PER CENT DISTILLED
JAH
5/23/51

FIG. 2
REFRACTIVE INDEX VS. B.P. @ 760 mm Hg





RECIPROCAL OF BOILING POINT IN °KELVIN X 10^3
SPECIFIC GRAVITY, 60° F/60° F
INDEX NUMBER
BENZENE
LEGEND
N- PARAFFINS
ISO-PARAFFINS
N-AND ISO-OLEFINS
CYCLOHEXANES
CYCLOHEXENES
TERPENES
SESQUITERPENES
MONOCYCLIC AROMATICS
CONDENSED AROMATIC NEUCLEI

REPORT OF CRUDE OIL ANALYSIS

OURCE Santa Cruz Tar Sands / LOCAT Santa Cruz, California DATE REC'D 3-4 SHIPPED ~~XXXXXX~~ N BY B. Shirley DATE REPORTED 3-5-57

ENERAL CHARACTERISTICS:

.P.I. @ 60°F.	24.5	Color	Black	Vis. SS U @ 77 F. =	63
P.Gr.@60°F.	.9071	Odor	Sour	Water by Dist.	
bs./Gal.	7.554	Sulphur		B.S. & W.	0.3
lash, °F.		Pour Point		Carbon Residue	
ire, °F.		Viscosity		Base of Crude	Naphthene

DISTILLATION, BUREAU OF MINES, HEMPLE METHOD

tmospheric Distillation Barometer 632 mm. First Drop 122 °F.

raction	Temp. °F.	Vol. %	Sum. %	Sp. Gr.	°API	C.I.	Vis. @ 100	Cloud °F.	
1	Up to 122	- - -	- - -						
2	122-167	.7	.7						
3	167-212	1.0	1.7						
4	212-257	1.3	3.0	.7783	50.3	37			
5	257-302	2.9	5.9	.7990	45.6	41			
6	302-347	5.3	11.2	.8198	41.1	43			
7	347-392	4.7	15.9	.8383	37.3	46			
8	392-437	9.5	25.4	.8545	34.1	48			
9	437-482	11.8	37.2	.8702	31.1	50			
10	482-527	12.1	49.3	.8855	28.3	53			
acuum Distillation @ 40 mm.									
11	Up to 392	8.7	58.0	.8996	25.8	56			
12	392-437	9.1	67.1	.9147	23.2	60			
13	437-482	8.1	75.2	.9315	20.4	64			
14	482-527	5.3	80.5	.9471	17.9	69			
15	527-572	4.8	85.3	.9646	15.2	75			
16		4.1	89.4	.9772	13.3	77			
17									
18									

'ESTS ON RESIDUUM:

Penetration @ 77°F.	6	Sp. Gravity @ 60°F.	
Viscosity @ 210°F. SFS		Conradson Carbon, %	
Flash, °F. coc		Pour, °F.	

UMMARY Residue 9.9 Vol. %

OREGON BASIN CRUDE

	E.P.	%	Gals.	@	Ext.
asoline	375	12	5.04		
aphtha	425	3	1.26		
D	510	7	2.94		
GO	650	10	4.20		
GO	850	21	8.82		
GO		3	1.26		
SP	8[illegible]	2	17.64		

CRUDE

	%	Gals.	@	Ext.
Gasoline	11.2			
Naphtha	4.7			
KD	21.3			
LGO	29.9			
HGO	18.2			
VGO	4.1			
ASP	9.9			
Loss	0.7			

Remarks: This analysis resembles the analysis on L-3 crude reported 55-W-39. Sulfurs were not determined on these cuts. Very aromatic material. Product discolor.

Freight Rate

Reported By J. Sheffield

1. Volymsändring vid pyrolys (7.1.52. GH)

En 800 ml Pyrexbägare av hög modell fylldes med tjärsand. I mitten nedsattes ett 300 watts el-element. Tjärsanden upphettades, tills all synlig gasutveckling upphört. Höjden av tjärsanden i bägaren steg under upphettningen från 133 till 142 mm. Den senare höjden kvarstod även sedan tjärsanden kallnat.
Runt elementet fanns efteråt en hård kokskärna, tjockare nedtill, uppåt avsmalnande. Utanför kärnan var sanden urlakad av oljeångorna och låg lös i kärlet. Denna sand gav en glödgningsförlust av 0.89%.

2. Vattenångdestillation (19.2.52. BP.)

100 g tjärsand och 200 g vatten satsades i en 1000 ml, trehalsad kolv, som placerades i en el. ugn och upphettades. Mellersta halsen försågs med uppsats, termometer, liebigkylare och förlag. Efterhand som vatten destillerade över, tillfördes nytt vatten via en skiljetratt i ena sidohalsen. Andra sidohalsen var proppad.
Efter 6 timmars upphettning hade i förlaget erhållits ca. 1500 ml vatten och blott 0,8 ml olja.

Slutsats: Vattenångdestillationen spelar förmodligen ingen roll, så länge blott opyrolyserad tjära föreligger

Undersökning av tjärsand
från
Santa Cruz, Californien.

Ett mindre prov av tjärsand från Santa Cruz, Californien, skickades till SSAB för att här skulle undersökas, om LINS-metoden kan provas i detta tjärsandsområde.

Tjärsandsprovet bestod av sandkorn omgivna av "tjära", som hade en kornig men fast och ganska hård struktur. Provet liknade tjärsanden från Alberta, Canada, men då det innehöll lägre halt "tjära" än Canada-sanden, var färgen ljusare, ungefär gråsvart, och sanden var ej plastisk utan betydligt hårdar. Vid uppvärmning blev dock tjärsanden så mjuk, att den kunde formas med handen.

Nedan angivna analysdata utom fukthalt och volymvikt av tjärsand är angivna på torrt prov. För jämförelse med tjärsanden från Alberta har en del analyser, som utförts vid SSAB, medtagits.

Tjärsand.	Santa Cruz	Alberta
Fukthalt, vikts-%	0,45	1,5
Volymvikt, g/cm³	1,88	1,90
Extraktion med tri-kloretylen, vikts-% "tjära"	13,0	18,1
Specifik vikt av extraherad sand, g/cm³	2,15	2,44
Siktanalys av extraherad sand, vikts-%		
Storlek, mm		
> 2		3,0
0,75 - 2	2,2	0,9
0,5 - 0,75		1,0
0,25 - 0,5		13,5
0,25 - 0,75	41,0	(14,5)
0,125 - 0,25	54,8	70,0
< 0,125	2,0	11,6
	100,0	100,0
Porositet, beräknad volym-%	0	0,6
Glödgningsförlust av tjärsand, vikts-%	14,3	19,7
C-total, vikts-%	10,43	13,9
C-karbonat, vikts-%	0,09	
H, vikts-%	1,36	1,7
S, vikts-%	0,89	0,9
Värmevärde, kal., kcal/kg	1.280	1.700

Fischer-pyrolys

	Santa Cruz	Alberta
Pyrolysvatten, vikts-%	0,7	0,4
Olja, vikts-%	8,2	11,2
Gas, vikts-%, Nl/kg inom parentes	1,3 (13,5)	1,3 (14,6
Koks	89,8	87,1
	100,0	100,0
Utbyte i % av "tjära"	63,0	62,0

Standardpyrolys.

3,90 kg tjärsand pyrolyserades på vanligt sätt till 535° under 11 tim.

Nedanstående produkter, omräknade till torr tjärsand, erhölls:

Pyrolysvatten,	5,4 ml/kg	0,6 vikts-%
Olja	80,4 "	7,2 "
Gas	15,2 Nl/kg	1,4 "
Koks	908 g/kg	90,8 "
		100,0
Utbyte i % av tjära		55,4
" " % " olja enl. Fischer		87,9

På grund av att temperaturinstrumentet ej fungerade tillfredsställande, kunde en jämn temperaturstegring ej erhållas. Likaså måste försöket avbrytas vid 535°. Temperaturen och pyrolysproduktionen som funktion av pyrolystiden framgår av diagram 1.

Pyrolysvatten

Ammoniak, g/l	4,9
Fenol, g/l	0,48
Specifik vikt, d_4^{20}	1,03

Olja

Spec. vikt, d_4^{20}	0,895
Brytningsindex, n_D^{20}	1,501
Pour point, °C	-10
Viskositet vid +20° C, cSt	7,9
+50° C, "	3,6
Bromtal	56
C, vikts-%	84,4
H, vikts-%	11,7
H/C, atom/atom	1,65
S, vikts-%	2,46
Värmevärde, kal., kcal/kg	10.220

Oljens specifika vikt och brytningsindex som funktion av oljemängden i ml/kg framgår av diagram 3.

ASTM-destillationen av totaloljan framgår av diagram 4.

Gas

H_2S, vol.-%	1,0	
CO_2	1,0	
CO	0,4	
H_2	22,4	
N_2	13,7	
C_nH_{2n}	6,7	61,5
C_nH_{2n+2}	54,8	
	100,0	

Kolvätena utgjordes av:

CH_4, vol.-%	37,6	
C_2H_4	2,2	
C_2H_6	9,1	
C_3H_6	2,8	
C_3H_8	4,0	10,0 vol.-% gasol
i-C_4H_{10}	0,4	
C_4 övriga	2,8	
C_{5+}	2,6	
	61,5	

Värmevärde, kal., (beräknat) kcal/Nm^3	9.710
Spec. vikt, g/Nl	0,954

Gasens sammansättning under försöket i vol.-% som funktion av gasmängden i Nl/kg visas i diagram 2.

Koks

C-totalt, vikts-%	4,25
C-karbonat, vikts-%	0,06
H, vikts-%	0,23
S, vikts-%	0,36
Glödgningsförlust, vikts-%	5,04
Värmevärde, kal., kcal/kg	310
Sintringstemperatur, °C	> 1.000

Koksen var hård och porös och j klibbig utom en mindre del i retortens botten.

Värmebalans vid pyrolys av 1 kg torr tjärsand.

Ingående kcal	1.280
Utgående kcal	
720 g olja	740
15,2 Nl gas	150
908 g koks	280
	1.170
Oredovisat	110
	1.280

Material- ~~Elementar~~balans vid pyrolys av 1 kg torr tjärsand.

	Svavel Utgående g/kg	Väte Utgående g/kg	Kol Utgående g/kg
Ingående, g/kg	8,9	13,6	104,3
Pyrolysvatten		0,6	
Olja	1,8	8,4	60,7
Gas	2,2	2,5	5,1
Koks	3,3	2,1	38,5
Oredovisat	1,6	0,0	0,0
Summa utg. g/kg	8,9	13,6	104,3

	S	H	C
i oljan	1,8 g	8,4 g	60,7 g
i gas	2,2	2,5	5,1
i koks	3,3	2,1	38,5
i p-vatten	–	0,6	–
oredovisat	1,6	0,0	0,0
Summa	8,9 g	13,6 g	104,3 g

Kommentarer och slutsatser.

Den undersökta tjärsanden från Santa Cruz liknar tjärsanden från Alberta utom beträffande tjärhalt och därmed sammanhängande faktorer, såsom erhållen mängd olja vid Fischer-pyrolys. Tjärsanden från Santa Cruz är ju betydligt "magrare" men ger dock tillräckligt med olja för att försöken med LINS-metoden skall kunna utföras, ehuru provet innehåller mer "tjära" än genomsnittet av tjärsandsfyndigheten i Santa Cruz, som enligt litteraturen är cirka 10 - 12 vikts-% mot 13 vikts-% i det undersökta provet.

Pyrolys av tjärsand är ej undersökt i samma utsträckning som pyrolys av skiffer. Om man jämför resultaten från standardpyrolysen med motsvarande undersökning av skiffer från Kvarntorp, synes dock pyrolysen ske på liknande sätt, såsom beträffande att oljebildningshastigheten uppnår maximum vid cirka 400°, och att gasbildningshastigheten uppvisar två maxima vid cirka 410° och 440°. Däremot var den erhållna oljans specifika vikt under hela pyrolysen lägre än specifika vikten av oljan från Kvarntorps-skiffer. En egendomlighet är, att oljans specifika vikt sjönk mot slutet av pyrolysen. Detta kanske berodde på följande: Då de bildade pyrolysgaserna passerar i retorten genom icke färdigpyrolyserad tjärsand, extrahera dessa gaser en del "tjära". Vid slutet av pyrolysen finns emellertid icke kvar något extraherbart, organiskt material, varför vid slutpyrolysen endast krackning sker, och då den krackade oljan har lägre specifik vikt än en lösning av denna olja och "tjära", sjunker oljans specifika vikt.

Oljeutbytet i % av tjära var för Santa Cruz-provet 63 % och för Alberta-provet 62 %. Motsvarande gasutbyten var 10 respektive 7 %. Sammanlagda olje- och gasutbytena blev sålunda 73 respektive 69 %, alltså ingen större skillnad. Dessa värden är emellertid betydligt högre än motsvarande utbyten för skiffer beräknade på kerogenhalten. För Kvarntorps-skiffern är dessa värden cirka 20 - 25 % för olja och cirka 18 - 19 % för gas, alltså totalt cirka 38 - 44 %. Denna stora skillnad mellan tjärsand och skiffer beträffande olje- och gasutbyten kan bero på följande:

1. Tjäran bildades under en senare tidsperiod (Krita- till Devon-perioden) än kerogenet (Kambrium-Silur-tiden) och utgöres kanske av en petroleumoljerest, varför den ej är så komplicerat uppbyggd som kerogenet.

2. Tjärsandens väte-kol-förhållande är högre än skifferns.

3. Den ovan nämnda extraktionen av pyrolysgaser.

Bensinhalten i oljan var i förhållande till oljans specifika vikt förvånande låg.

Vid pyrolys av skiffer blir som bekant oljeutbytet lägre ju långsammare pyrolysen sker. Av resultaten från pyrolyserna enl. Fischer och "standardmetoden", där pyrolystiden var 1,5 respektive 10 tim. till 500° synes, att detta även gäller för tjärsand. Om detta emellertid gäller i samma utsträckning som för skiffer är emellertid omöjligt att säga. Vid exempelvis en pyrolys in situ, där pyrolystiden är lång, får troligen "extraktionsverkan" stor betydelse, varför det mycket väl kan tänkas, att oljeutbytet vid en pyrolystid av 0,5 - 1 år in situ ej blir nämnvärt lägre än vid ovanstående "standardpyrolys".

Vid jämförelse av gasutbytena vid Fischer- och "standard"-pyrolyserna ökade gasutbytet från 13,5 till 15,2 Nl/kg tjärsand vid den längre pyrolystiden, vilket överensstämmer med pyrolys av skiffer. Vid betydligt längre pyrolystider, som vid in-situ-pyrolys, sjunker emellertid gasutbytet åter. Detta blir troligen också gällande för tjärsand, särskilt om det antages, att en del tjära extraheras och alltså ej pyrolyseras. Detta är dystert med tanke på att det vid pyrolys enligt LINS-metoden är önskvärt, att den okondenserbara gasen skall kunna användas till brännarna och vara tillräcklig för upprätthållande av pyrolysen. Det fordras troligen cirka 175 kcal gas för att pyrolysera 1 kg tjärsand med LINS-metoden. Vid "standardpyrolysen" ovan erhölls emellertid endast 150 kcal gas vid pyrolys av 1 kg tjärsand.

Närkes Kvarntorp den 17 februari 1955.

Bengt Persson

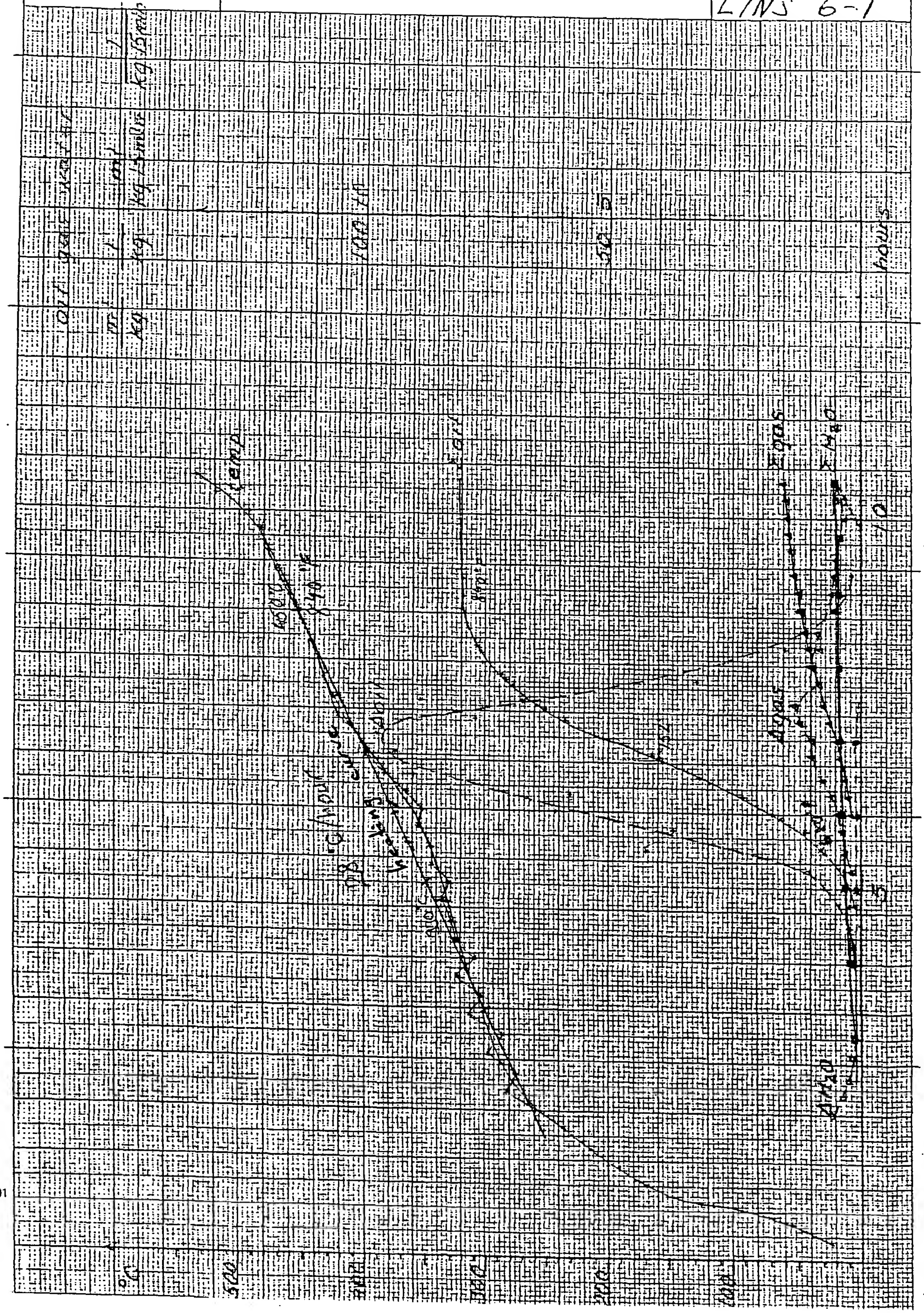
1 kg sand med sp. värme 0,25 cal/g. °C värmes från +15°C till 400°C:

$$\frac{1000 \cdot 0{,}25 \cdot 385}{1000} = \underline{\underline{96 \text{ kcal/kg.}}}$$

Standardpyrolysis

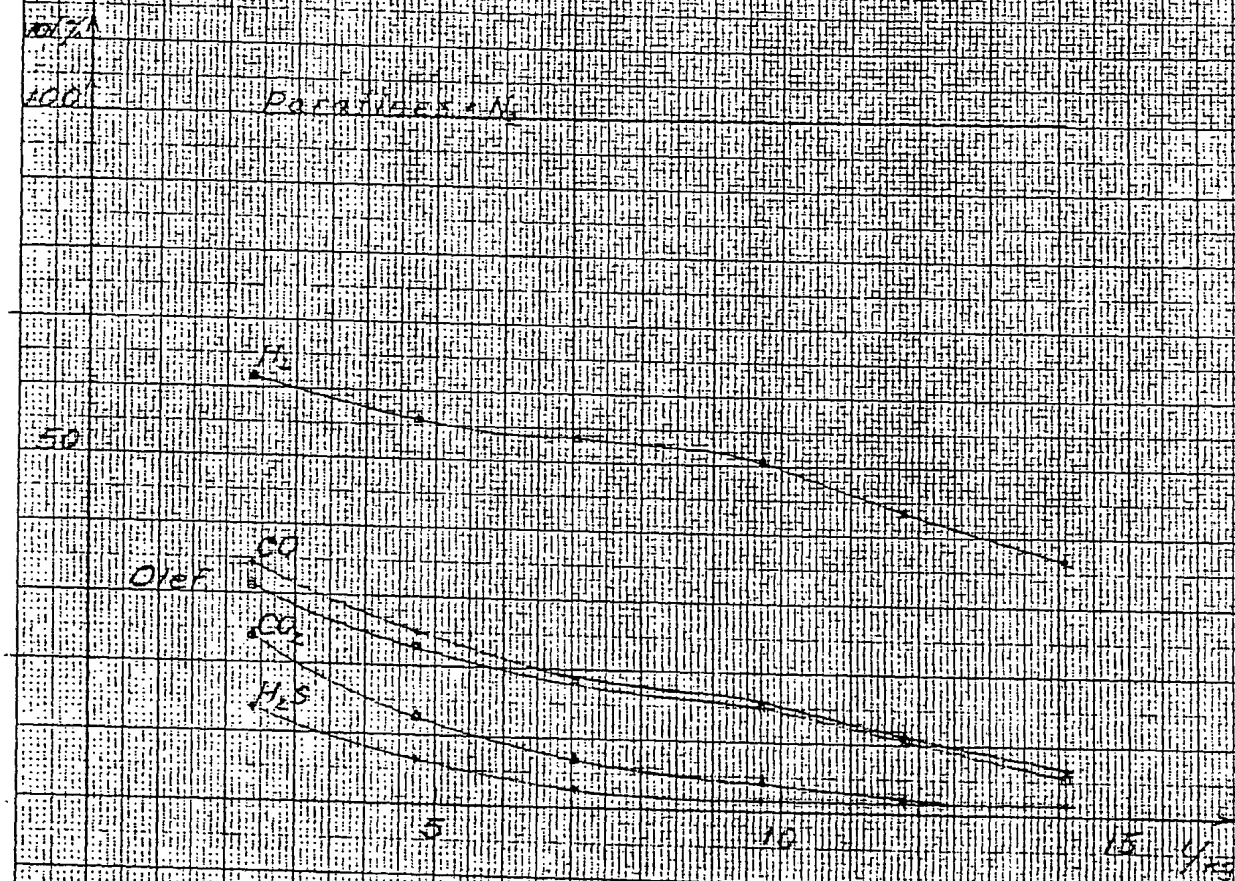
12.10.1954

LINS 6-1



Composition of crude gas | 12.10.19[illegible]

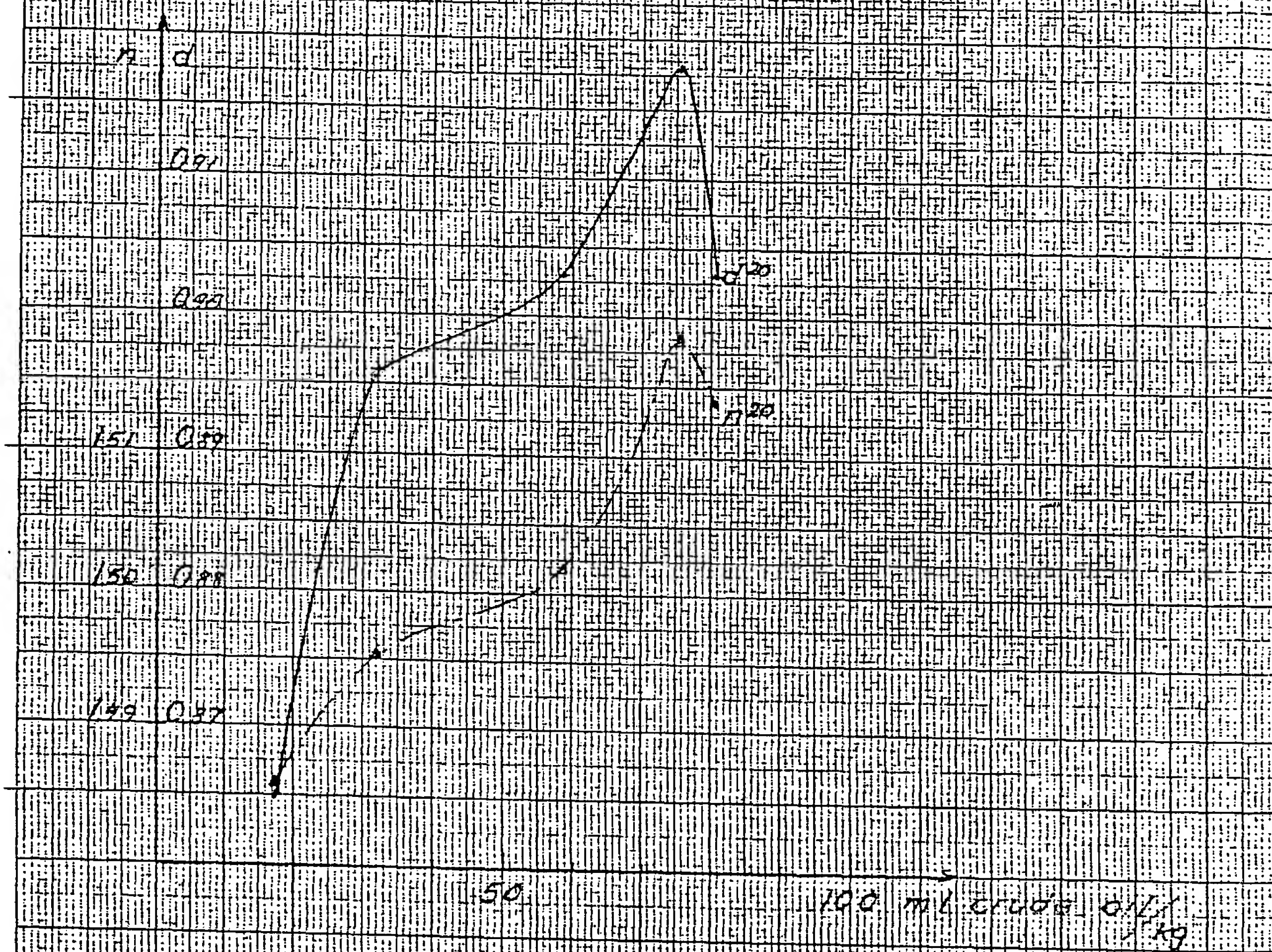
LINS 6-2



Refractive index and spec. gravity of pyrolysis oil

12.10.1954

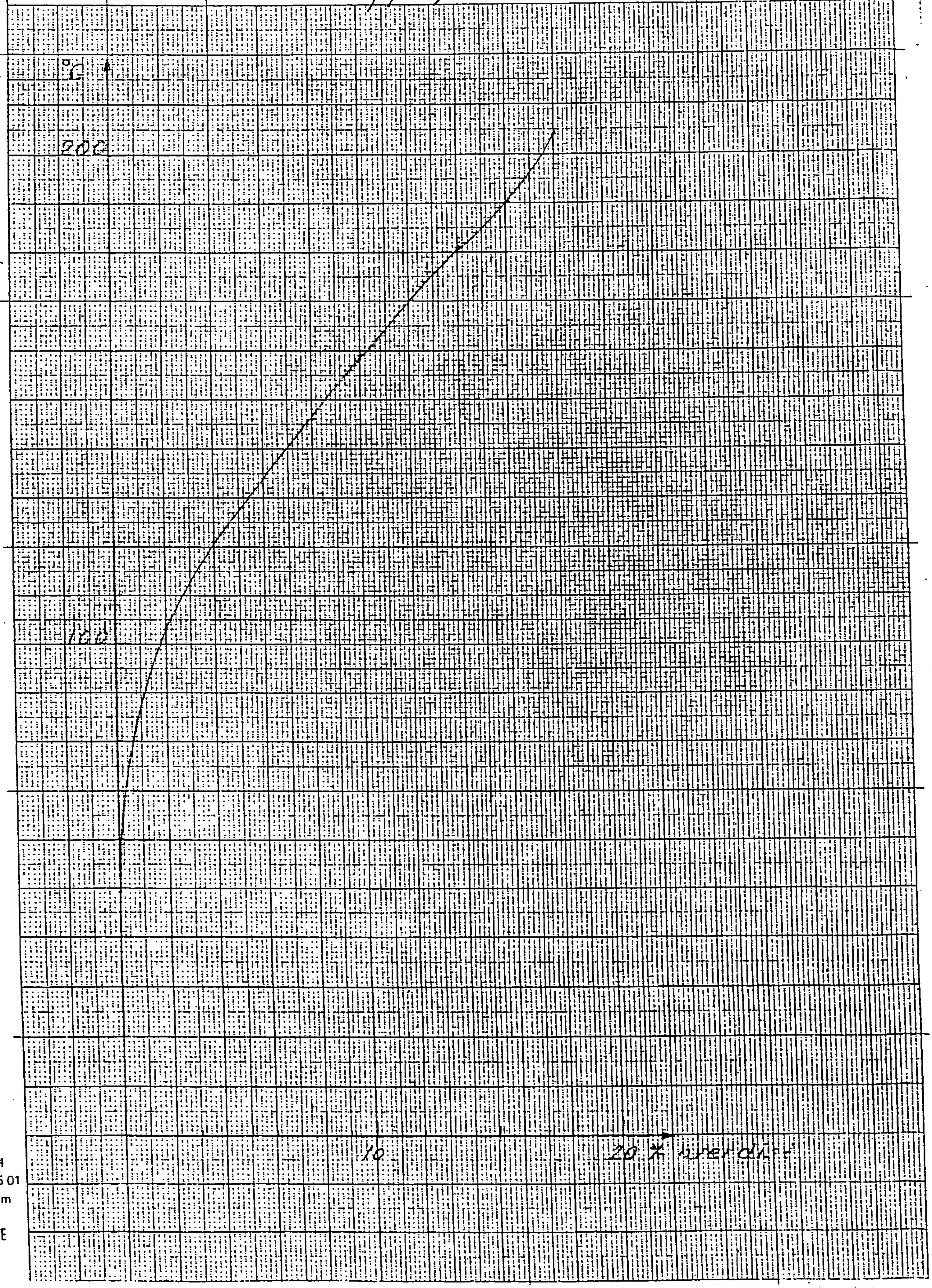
LINS. 6-3



23 A4
3 25 01
x1 mm
ESSELTE
4416

ASTM-distillation
of pyrolysis oil

12.10.1984
LINS 6-4



Undersökning av tjärsandspyrolys

Värmetransporten i kompakt tjärsand är relativt långsam. Då en viss temperatur (omkring 300° C) överskrides, inträder pyrolys, som ger upphov till oljeångor och gaser, som båda lämnar pyrolyszonen, samt koks, som kvarstannar i håligheterna mellan sandkornen. Formen av den sammanhängande kokskropp, som bildas, ger alltså en påtaglig bild av hur värmet spritt sig omkring ett värmeelement.

De bildade pyrolysångorna kondenseras delvis i kallare partier av tjärsanden. En del av tjäran löses i de kondenserade pyrolysprodukterna. Lösningens viskositet blir lägre än tjärans, och en viss strömning av olje-tjärablandningen kan väntas ske, exempelvis nedåt mot lägre liggande lager och naturligtvis under förutsättning att fri porvolym finnes. Även ren tjära kan naturligtvis flyta från en zon till en annan i samma mån som inträngande värme reducerar tjärans viskositet.

För studier av värme- och materialrörelserna i tjärsanden gjordes 1952 och 1953 ett antal modellförsök i laboratorieskala vid pyrolyslaboratoriet i Kvarntorp. För de mindre försöken användes därvid tjärsand från Athabasca, Canada. För vissa försök i större skala åtgick större kvantiteter tjärsand än vad som bekvämt kunde erhållas från Athabasca. Då det i dessa större försök ej var frågan om kvantitativa eller kvalitativa studier av de erhållna produkterna, ansågs det fullt tillfredsställande att för försöken använda en "syntetisk" tjärsand, tillverkad genom intim blandning av fin kvartssand med uppvärmd tjära i så nära samma egenskaper som Athabasca-tjärans som möjligt. En jämförelse mellan den genuina Athabasca-tjärsanden och den syntetiska dito finnes i bifogade analystabell. Ca 60 ton av den syntetiska tjärsanden fylldes i en kvadratisk låda med en ca fyra meters sida och ca två meters djup och packades tätt i varmt tillstånd. Ovan tjärsanden packades ca ett n meter tjockt lager pinnmo, avsett att motsvara den s.k. overburden, som finnes över naturliga tjärsandsförekomster. I mitten av lådan nedsattes sju elektriska värmeelement, inneslutna i 1½-tums vertikala järnrör och arrangerade i form av en sexkant med 1,5 meters kantlängd och med ett värmeelement i varje hörn och ett i centrum. Vidare upptogs tt antal mäthål, i vilka placerades termoelement, och ett antal provtagningshål för gas- och oljeångor.

Koncentriskt omkring värmeelementrören placerades perforerade gasrör för utsläppning av pyrolysprodukterna. Hålarrangemanget framgår även av bifogad skiss.

Värmeelementen inkopplades, och uppvärmningen fick pågå omkring en vecka, då det bedömdes, att pyrolysen hade fortskridit så långt, att en lätt studerbar värme- och materialfördelning erhållits i provlådan. Så snart lådan och dess innehåll svalnat tillräckligt mycket för att en närmare undersökning skulle kunna ske, uppgrävdes lådans innehåll försiktigt. Speciellt beaktades att ingen ändring av materialfördelningen mellan olika delar av tjärsanden åstadkommits genom själva utgrävningsarbetet. Ett stort antal prover uttogs och analyserades. Resultaten anges i bifogade tabeller.

Närkes Kvarntorp i mars 1958

Gösta Salomonsson

(Gösta Salomonsson)
Överingenjör

1. Produkternas rörelser i sanden.

Glödförlusten anger koks + olja + tjära. Differensen mellan den aktuella punktens glödförlust och den ursprungliga sandens (16,0 %) anger hur mycket organisk substans (olja + tjära), som netto strömmat till eller från punkten. Differensen betecknas A.

$$\therefore\ A = 16{,}0 - G$$

2. Pyrolysgraden. Tjäran och oljan kan ha förflyttat sig i sanden efter pyrolysen, men så är icke fallet med koksen. Mängden koks är proportionell mot mängden pyrolyserad tjära och utgör sålunda ett mått på pyrolysgraden. Koksmängden erhålles som differensen (=B) mellan glödningsförlusten och den extraherbara substansen.

$$\therefore\ B = G - E = \text{koks}$$

3. Eventuella urlakningseffekter.

En del tjära löses i oljan och medfölja denna på dess rörelser i sanden. En uppdelning av den extraherbara substansen (E) på tjära och olja är sålunda önskvärd. En dylik erhålles genom Fischeranalysen.

Fischerprovets oljeutbyte F = olja, bildad ur förhandenvarande tjära (x) i punkten + avdestillerad olja (y), som från annat håll runnit till punkten ifråga.

den förhandenvarande oljebildningen
Men ~~x~~ är prop. mot $x, = \alpha \cdot x$

$$\therefore\ F = \alpha \cdot x + y$$

Vidare är $E = x + y$

$1-\alpha$

Enligt Fischeranalysen av det ursprungliga fjärsandsprovet är $1-\alpha = 0,31$.

$$\therefore \quad x = 3,23\,(E-F) = \text{fjära}$$
$$y = E - x \quad = \text{olja}$$

Det bör här anmärkas, att differensen $E-F$ och alltså x är behäftad med stora fel.

De ur analyserna framräknade koks-, fjär- och olje-halterna anges i tabell 3.

Tabell 2. Analyser av prover, uttagna på olika ställen i tjärsandlådan

Avstånd från centrumelementet, cm →	0 (vid [illegible])			10			20			30			40			65			85		
Avstånd från lådans botten, cm ↓	G	F	E	G	F	E	G	F	E	G	F	E	G	F	E	G	F	E	G	F	E
(täcksanden)	—	—	—	—	—	—	—	—	—	—	—	—	5,8	2,2	—	—	—	—	—	—	—
150	—	—	—	20,2	20,4	0,3	17,7	21,0	0	2,3	20,0	0,3	3,1	17,0		3,7	20,9		3,0	19,0	
100	—	—	—	17,6	18,9	20,0	18,3	20,6	0	20,1	20,0	21,4	16,9	7,0		17,1	17,0		10,2	18,1	
50	7,4	1,9	—	20,5	0	21,4	22,3	0,2	0	20,5	0,4	18,6	20,8	0,1		20,3	3,3		10,9	2,5	
0	22,0	15,5	—	20,1	13,5	—				18,8	13,2	0									

G = glödgningsförlust = koks + tjära + olja
F = oljeutbyte vid Fischerprov
E = extraherbart material = olja + tjära

Tabell 3. Koks-, tjära- och oljehalter, %.

Avst. fr. centrum, cm →	0 (rörväggen)			10			20			30			40			65			85		
Avst. fr. botten, cm ↓	koks	olja	tjära	koks	olja	tjära	koks	olja	tjära	koks	olja	tjära	koks	olja	tjära	koks	olja	tjära	koks	olja	tjära
(täcksanden)																					
150				19,5	0	0,6				2,0	0	0,6									
100				0	0	20,3				0	0	21,3									
50				0	0	21,7				1,9	0	20,7									
0																					

(Anm. De oljehalter, som betecknats med 0, motsvara vid uträkningen enl. ovan, små negativa differenser y=E−x. Inom försöksfelen kan det anses att ingen olja, endast tjära fanns på dessa punkter.)

Laboratorieförsök

	Tjärsand från Athabasca	"Syntetisk" tjärsand	T.s. från Santa Ana
Volymvikt, g/cm³	1,9	1,9	1.63 x)
Vattenhalt, %	1,5	0	0.44
Glödgningsförlust, %	19.7	16.0	14.25
Porositet %	2	6	
Fischerpyrolys			
vatten %	0,4	1,7	0,7
olja, %	11,2	11,0	8,2
gas Nl/kg	14,6	11,3	13,5
koks, %	87,1	86,3	89.8
Permeabilitet, millidarcys	2	200	
Siktanalys, glödgad sand %:			
siktstorlek, 2,0 mm	3,0	0,0	
1,0	0,9	0,1	
0,5	1,0	0,3	
0,250	13,5	24,2	
0,125	70,0	74,4	
0,074	7,3	0,6	
<0,074	4,3	0,1	
Värmevärden, Mcal/kg: tjära	9.8-9.9	9.85	
— " — tjärsanden	1,70	1.60	1.28
Spec. vikt: tjära	1.02	1.03	1.097?
Elm-Extraktion tjära i prov, %	—	—	13.2

x) extraherad sand volym- = ...

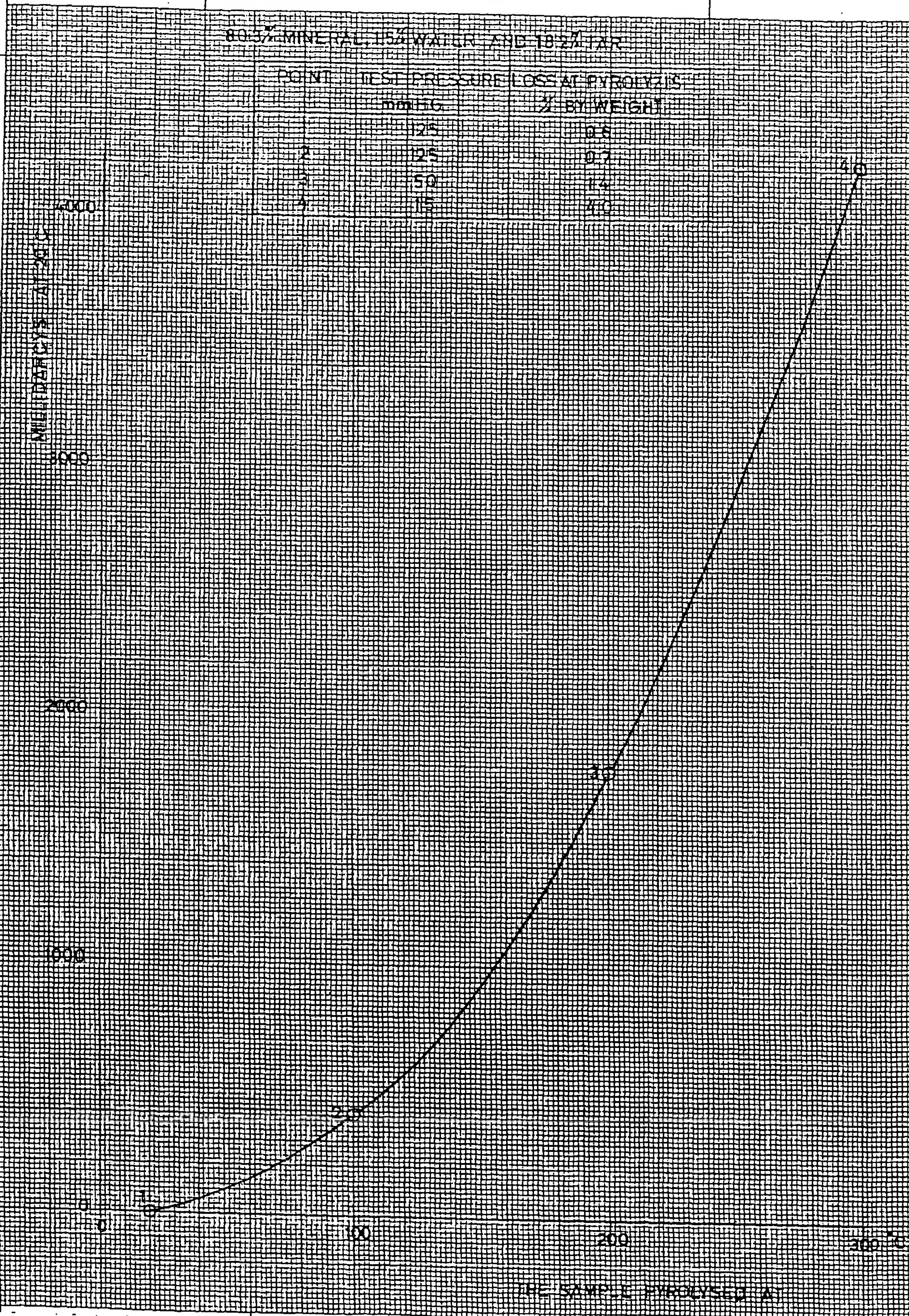
S.S.A.B.

~~At~~ heated to VARIOUS TEMPERATURES.

BIL 11

80.5% MINERAL, 1.5% WATER AND 18.2% TAR

POINT	TEST PRESSURE mmHG	LOSS AT PYROLYZIS % BY WEIGHT
1	125	0.6
2	125	0.7
3	50	1.4
4	15	4.0



523 A 4
1 mm
ESSELTE
4446